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# Carver-Greenfield Process Evaluation; A Process for Sludge Drying

Los Angeles-Orange County Metropolitan Area Project, Whittier, CA

Prepared for

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The Carver-Greenfield process, which operates on the principle of multiple-effect evaporation, offers a considerable potential for drying wastewater solids. Drying of dewatered sludge cake by this method before incineration or pyrolysis offers a potential for energy self-sufficiency for the total sludge handling system. A continuous flow pilot-scale demonstration of the Carver-Greenfield process was conducted by the LA/OMA Project. A four month long demonstration study was conducted by the City of Los Angeles at their Hyperion Treatment Plant under a contract from the LA/OMA Project. The LA/OMA Project staff also visited an operating Carver-Greenfield installation treating waste activated sludge (Coors Brewery, Golden, Colorado) and another defunct installation which was plagued by operating problems (Omaha, Nebraska). From the pilot-scale demonstration study as well as from the field trips it was felt that the Carver-Greenfield process is a viable means for drying wastewater solids. The system did not seem to be too sophisticated for the use in a sewage treatmen, ....

KEYWORDS: *Sewage sludges, *Carver - Greenfield process,
*Sludge treatment.

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ABSTRACT

## CARVER-GREENFIELD PROCESS EVALUATION

The LA/OMA Project is to develop a long-term plan, including an implementation strategy to re-use/dispose of residual solids from industrial and municipal wastewater treatment in the Los Angeles - Orange County metropolitan area.

This report presents a study on the Carver-Greenfield process, a multiple-effect evaporation system which offers considerable potential for drying wastewater solids. Under a contract from the LA/OMA project, a demonstration study was conducted by the City of Los Angeles treatment plant. It was felt from this investigation, that a large demonstration plant is needed to collect reliable design and operating data.

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LA/OMA PROJECT

# CARVER - GREENFIELD PROCESS EVALUATION

## A PROCESS FOR SLUDGE DRYING

DECEMBER 1978

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# CARVER-GREENFIELD PROCESS EVALUATION

## A PROCESS FOR SLUDGE DRYING

**Participating Agencies**
U.S. Environmental Protection Agency (Region IX)
California State Water Resources Control Board
City of Los Angeles
Los Angeles County Sanitation Districts
Orange County Sanitation Districts

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# FOREWORD

The Carver-Greenfield process, which operates on the principle of multiple-effect evaporation, offers a considerable potential for drying wastewater solids. In a conventional drying process more than 1000 BTU of thermal energy is needed to evaporate 1 pound of water. But in a multiple-effect evaporation process, for example, in a four-effect system, the theoretical thermal energy requirement is only 250 BTU/pound of water. Although, considering for various losses and heating efficiency, the actual heat requirement for such a system is about 430 pounds of steam per of water. Drying of dewatered sludge cake by this method before incineration or pyrolysis offers a potential for energy self-sufficiency for the total sludge handling system. Although the Carver-Greenfield process has been practiced in the food processing industry for a considerable number of years, it is a relatively recent development in the traditional sewage treatment operation.

In view of the above, a continuous flow pilot-scale demonstration of the Carver-Greenfield process was conducted by the LA/OMA Project. A four month long demonstration study was conducted by the City of Los Angeles at their Hyperion Treatment Plant under a contract from the LA/OMA Project. The LA/OMA Project staff also visited an operating Carver-Greenfield installation treating waste activated sludge (Coors Brewery, Golden, Colorado) and another defunct installation which was plagued by operating problems (Omaha, Nebraska).

From the pilot-scale demonstration study as well as from the field trips it was felt that the Carver-Greenfield process is a viable means for drying wastewater solids. The system did not seem to be too sophisticated for the use in a sewage treatment plant. Unless properly designed the system could loose considerable amount of process oil, emit hydrocarbons to the atmosphere and create odor problem. On the basis of the experience gained from the present investigation, it was felt that a moderately large-scale demonstration plant is needed to collect reliable design and operating data.

# REGIONAL WASTEWATER SOLIDS MANAGEMENT PROGRAM

# LOS ANGELES/ORANGE COUNTY METROPOLITAN AREA

## LA/OMA PROJECT

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# CONCLUSIONS

# CONCLUSIONS

1. The Carver-Greenfield system appears to be a viable sludge drying process and offers considerable energy efficiency compared to conventional direct or indirect contact dryers. However, a large scale facility should be built and operated to conclusively demonstrate process reliability and economics.

2. Technology involved in the Carver-Greenfield process is not beyond that which can reasonably be expected to be operated in moderate to large treatment plants.

3. Energy requirement for a four-effect Carver-Greenfield system with hydroextraction is projected to be about 0.44 lb of steam per lb. $H_2O$ evaporated. This value is based on data supplied by the manufacturer, data determined for the Coors facility, and supported by theoretical analysis of the system. Assuming a boiler efficiency of 75 percent, and a heat requirement of 1150 BTU/lb of 65 psia saturated steam produced, the energy requirement including steam production would be about 675 BTU/lb $H_2O$ evaporated. This compares favorably with the 1200 to 2000 BTU/lb $H_2O$ required in most conventional heat dryers.

4. Even though the C-G system is energy efficient compared to conventional dryers, energy requirements can still be significant. Therefore mechanical dewatering prior to the Carver-Greenfield is highly advised to reduce the evaporative burden.

5. The process is capable of producing 98 percent dry solids. If pyrolyzed or incinerated, such solids would support autogenous combustion without the need for continuous supplemental fuel addition. Carver-Greenfield drying of raw waste activated sludge (WAS) and blending with digested, dewatered primary sludge (PS) can also produce an autogenous mixture, depending on the WAS/PS ratio and the characteristics of the two sludges.

6. Thermal processing of a mechanically dewatered, C-G dried product, by incineration or pyrolysis processes, can produce sufficient steam and electrical power to make the C-G-thermal processing system energy self-sufficient, and can even produce surplus energy.

7. For the case of Carver-Greenfield drying of combined dewatered sludge, and under conditions examined in this report, about 35 to 45 percent of the steam produced during thermal processing of the dried product would be needed for operation of the Carver-Greenfield system, with the remainder available for electrical power production.

8. A comparison was made between Carver-Greenfield drying and thermal processing of a combined sludge with copyrolysis of dewatered cake using refuse derived fuel (RDF) as a fuel supplement. Assuming 390 dtpd of digested, dewatered combined sludge and other conditions as described in the report, about 130,000 KWH/day net electricity would be produced from the CG dried product. Addition of about 421 tpd RDF to achieve autogenous combustion of a dewatered cake would yield about 87,000 KWH/day net electrical production. Increasing the RDF to 590 tpd increases the net electrical output to about 158,000 KWH/day for the case of dewatered cake.

9. Under the conditions assumed, C-G drying of dewatered cake followed by thermal processing with energy recovery in a multiple hearth furnace presented a lower net cost per ton of sludge solids than the co-pyrolysis cases examined.

10. The process condensate was found to be highly biodegradable. Essentially all of the sludge feed water, except for small amounts in the final product, will appear as process condensate. COD concentrations ranged from about 300 to 2500 mg/l. For liquid and dewatered digested sludges, about 0.008 to 0.012 lb COD distilled into the condensate per lb. of dry sludge fed. For raw WAS about 0.08 lb. COD distilled into the condensate per lb. of feed solids.

11. Process condensate was found to contain 80-95 percent of the $NH_3$-N contained in the feed sludge. No significant decomposition or distillation of organic-nitrogen was observed.

12. Oxygen demand, imposed on secondary treatment facilities from recycle of the process condensate, is projected to average less than 5 percent of that exerted by the primary effluent flow in the case of dehydration of liquid sludge, and less than 1 to 2 percent in the case of dewatered sludge. No other constituents which would likely impact operation of the treatment plant were found in the condensate. However, the effect of ammonia should be considered on processes such as nitrification and breakpoint chlorination. It is concluded that the process condensate could be directly recycled without producing significant impact on the treatment plant.

13. Additional fluidizing oil recovery, beyond that achieved by centrifugal dewatering, is essential to make the process economically viable and reduce dependence on outside sources of fluidizing oil. Hydroextraction of the centrifuged oil-solids mixture was shown to be effective in recovery of the remaining oil. Fluidizing oil losses in the hydroextracted product was about 1.0 percent or less of the infeed solids by weight.

14. There are multiple sources for potential loss of process oil in vapor form, particularly if Isopar is used as a carrying fluid. These include vacuum pump discharge, fluidizing oil storage facilities, and solids dewatering and conveyance facilities. All such facilities will need to be covered and/or vented to collect the oil vapors. The vapors should be easily cooled and condensed with remaining gases passed to the boiler or pyrolysis facility for combustion of any residual hydrocarbons. However, unless such vapor losses are controlled through proper design and operation, the system could create a financial burden, air pollution, and odor problems.

15. No scaling or clogging of heat exchanger tubes was noted during pilot testing. Fibrous material, such as may be contained in primary sludge, will require disintegration or grinding prior to being introduced into the system. Waste activated sludge may not require the disintegration step.

16. Fluidizing oil will tend to extract oils and greases contained in the sludge solids. Buildup of fine sludge solids can also occur. Steam stripping of a portion of the recirculating oil may be required to recover the extracted heavy oils and reducing the buildup of fine sludge solids.

17. Based on limited data, the Carver-Greenfield process appears to be effective in the destruction of pathogenic microorganisms.

18. While the present study has answered many questions regarding application of the Carver-Greenfield process to municipal sludges, some concerns and areas requiring further study still remain. These include:

    a. Further documentation on the mass and energy balance of a CG system employing such features as hydroextraction, heavy oil separation, and solids addback.
    b. Potential problems related to gummy phase formation and the effectiveness of solids addback as a control measure.
    c. Potential problems related to buildup of fine solids in the system during long-term, continuous operation.
    d. Potential scaling and corrosion problems during long-term, continuous operation.
    e. Effectiveness of containment and collection systems for vaporized fluidizing oil.
    f. Suitability of the dry product for feed to a thermal process and operating difficulties associated with the thermal process.

    Available information suggests that these concerns can be overcome through proper design and operation of the process. However, it is unlikely that remaining questions and areas of concern will be fully answered until such time as a large scale facility is designed and operated on municipal sludge.

Additional avenues for information development on the Carver-Greenfield process which might be pursued include; (1) site visits to the two Japanese plants operation on night soil; (2) further study of the Coors facility.

19. Partial slagging and fusion of pelletized Carver-Greenfield dried sludge was observed during pyrolysis tests at 1850°F in a vertical shaft furnace. Reducing the temperature to 1500°F reduced but did not eliminate the problem of fusion.

20. Carver-Greenfield dried sludge was successfully pyrolyzed in a rotary hearth furnace. No slagging or other material handling problems were observed at a temperature of 1500°F. Pelletizing of the dried sludge may not be required in this type of reactor. Results are thought to be applicable to multiple hearth furnaces as well.

21. Hearth loadings of 7.5 $\text{lb/hr-ft}^2$ were found to be suitable in the demonstration tests. Bench scale testing indicated that higher loadings may be possible. A value of 10 $\text{lb/hr-ft}^2$ was assumed for purposes of cost analysis.

22. Hydroextracted solids had a powdery consistency and might create a dust problem in conveyance, storage and hauling unless properly handled. Consideration must also be given to avoiding potential explosion hazards posed from dusting of the dry solids. Pelletizing may be necessary in some cases depending on use of the final product.

## RECOMMENDATIONS

On the basis of the experience gained from the pilot plant operations, field visits, and theoretical analyses conducted on the Carver-Greenfield process, and since it is believed that further pilot-scale studies would not generate substantial additional information on system reliability, a large-scale demonstration Carver-Greenfield plant of sufficient capacity should be undertaken to collect reliable design and operating data. Such data would be extremely helpful in determining the suitability of this process for use in large wastewater treatment plants.

# CHAPTER 1

# CHAPTER 1

## INTRODUCTION

A dehydration process called Carver-Greenfield process (patented by Dehydrotech Corporation, formerly Carver-Greenfield Corporation) has been applied to dry various type of organic and inorganic slurries in different industrial processes, including paper mills and food-processing plants, as well as some industrial wastewater solids. The possibility exists that this technique can successfully be applied to dehydrate sewage sludges.

In the Carver-Greenfield system, water is extracted from sludge by evaporation using a multi-effect evaporator. A flow diagram describing how the process fits in a total sludge management system is shown in Figure 1.1. Sludge to be processed is first thickened or dewatered as much as possible so as to reduce the amount of water to be evaporated. Thickened sludge is then mixed with an oil (carrying medium), such as No. 2 fuel oil or Isopar L (an Exxon product) at a suggested ratio of 1 part dry solids to 5 to 10 parts oil (1,2). By use of an oil, fluidity is maintained in all effects of the evaporation cycle and formation of scale or corrosion of the heat exchangers is minimized. The sludge-oil slurry is then pumped to the multi-effect evaporator where water is vaporized. The remaining solids-oil mixture is subsequently centrifuged to separate the oil and solids. The oil is recycled and reused while the solids are discharged for further processing or disposal.

Multi-effect evaporation affords an economy of steam over single-effect operations through the reuse of heat. The Carver-Greenfield process employs reverse flow, multi-effect evaporation with steam being added to the last effect. For example, in a three effect system, vapor from the last effect is used to provide heat in the second effect and the vapor from the second furnishes heat to the first. Vapor from the first effect is removed, condensed, and discharged. Through the reuse of heat in the multiple-effect process, the amount of water removed per pound of steam supplied increases with increasing number of effects.

In its simplest theoretical form, a single-effect evaporator can evaporate a maximum of one pound of water per pound of steam supplied and a two effect evaporator will evaporate two pounds of water per pound of steam supplied, because of the reuse of heat. Depending on the number of effects used, the amount of heat energy required per pound of water removed will vary. For a single-effect unit, at least 1000 BTU per pound of water removed is required and for a double-effect unit, at least 500 BTU is required per pound of water removed, etc. Vacuum is applied to the various effects to produce a temperature difference ( T) between effects and thus allow heat transfer between effects.

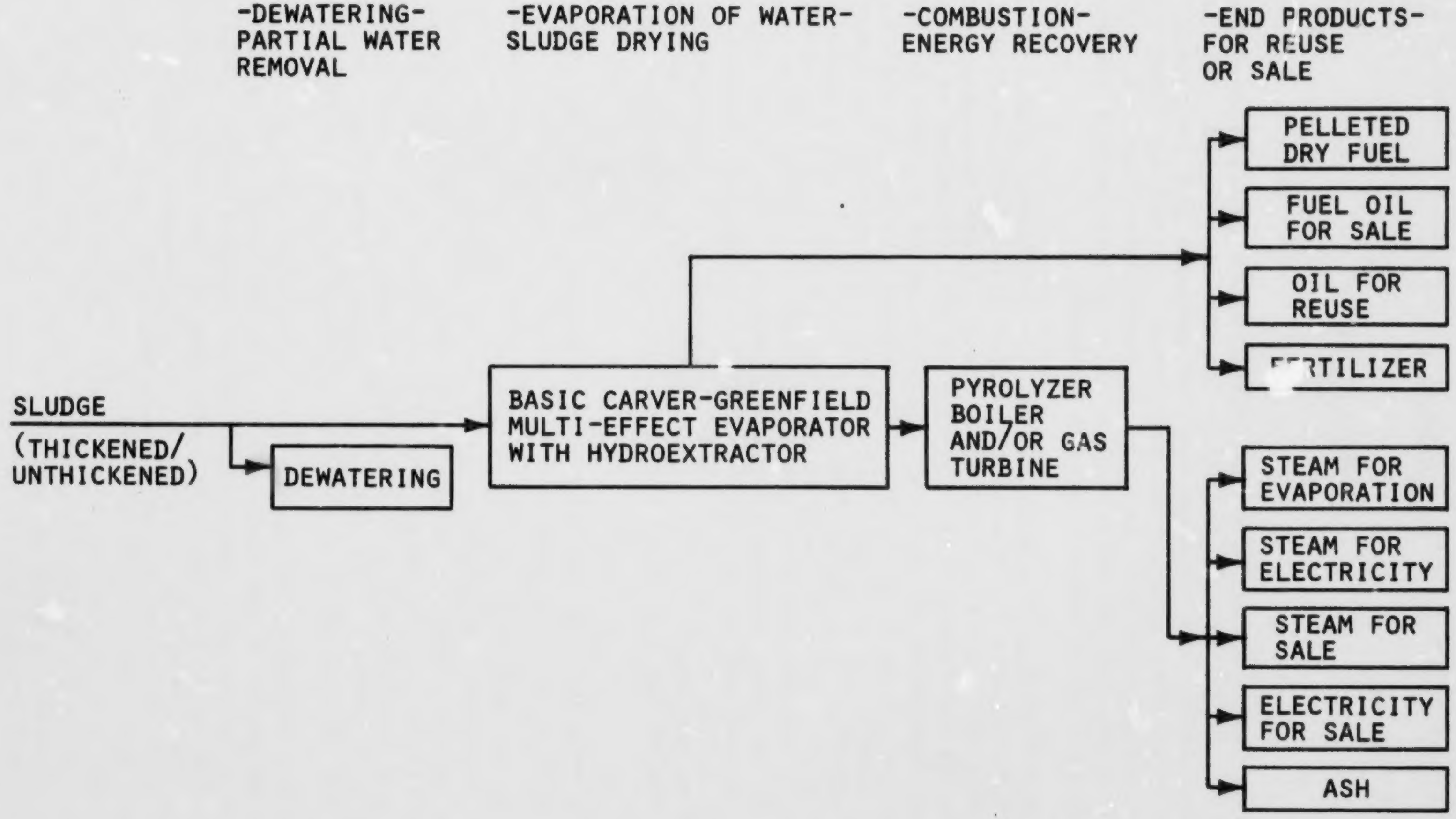

-DEWATERING-
PARTIAL WATER
REMOVAL
-EVAPORATION OF WATER-
SLUDGE DRYING
-COMBUSTION-
ENERGY RECOVERY
-END PRODUCTS-
FOR REUSE
OR SALE
SLUDGE
(THICKENED/
UNTHICKENED)
DEWATERING
BASIC CARVER-GREENFIELD
MULTI-EFFECT EVAPORATOR
WITH HYDROEXTRACTOR
PYROLYZER
BOILER
AND/OR GAS
TURBINE
PELLETED
DRY FUEL
FUEL OIL
FOR SALE
OIL FOR
REUSE
FERTILIZER
STEAM FOR
EVAPORATION
STEAM FOR
ELECTRICITY
STEAM FOR
SALE
ELECTRICITY
FOR SALE
ASH

**FIGURE 1.1**

FLOW DIAGRAM ILLUSTRATING THE POTENTIAL OF CARVER-GREENFIELD PROCESS IN WASTEWATER SOLIDS MANAGEMENT SYSTEM

LA/OMA PROJECT

It would appear that an infinite economy of steam would result from the use of an infinite number of effects. Several factors, however, limit the number of effects practicable in a system. Each effect of a multiple-effect evaporator operates only on a fraction of the total temperature drop across the system. The total drop is seldom larger than that employed in single-effect evaporation, and the capacity per unit area of heating surface is reduced proportionately. Thus, a savings in fuel requirements may be realized through multiple-effect operation but equipment costs will be greater. In most cases, no more than three or four effects are economical but the actual number is largely influenced by prevailing fuel costs.

Most proposals on treating municipal sludge with the Carver-Greenfield process include an incinerator or pyrolysis reactor to recover the heat value of the dried product. Theoretically, this is an attractive combination of processes since water can be evaporated with multiple-effect efficiency prior to combustion or gasification. Fuel gases produced during pyrolysis or waste heat from an incinerator can then be used to supply the energy requirements of the Carver-Greenfield process.

If energy production is desired, preconcentration of the sludge by centrifugation or filtration prior to evaporation is attractive since the quantity of water to be evaporated is greatly reduced as shown in Figure 1.2. With less water to evaporate, more net energy can be produced from the multiple-effect evaporation-combustion process. The manufacturer's estimation of net energy production is presented in Figure 1.3 for sludges of different heat value. A pyrolysis reactor was assumed in the analysis with resultant fuel gases used to produce steam at 85 percent thermal efficiency. A steam turbine was then used for electrical generation. Thermal efficiency of the steam turbine was not quoted. The number of effects considered economical by the manufacturer over a range of sludge concentrations is also presented in Figure 1.3.

## STUDY BACKGROUND

Although there are many Carver-Greenfield (C-G) installations in operation, they are mostly used in food processing industry or similar industrial operations. Included in these are several sludges from industrial waste treatment which may have characteristics similar to municipal sewage sludge. There are two installations in Japan which treat night soil. Recently, the Adolph Coors Brewery at Golden, Colorado has built a Carver-Greenfield plant to process waste activated sludge produced from their brewery treatment operation.

The C-G process is not a traditional sewage sludge handling practice, and concerns have been raised about the adaptability of the Carver-Greenfield process in municipal sludge treatment operation. There are also specific areas of concern about this process raised in the recent years. The most frequently mentioned concerns include the potential for the heat exchanger tube clogging

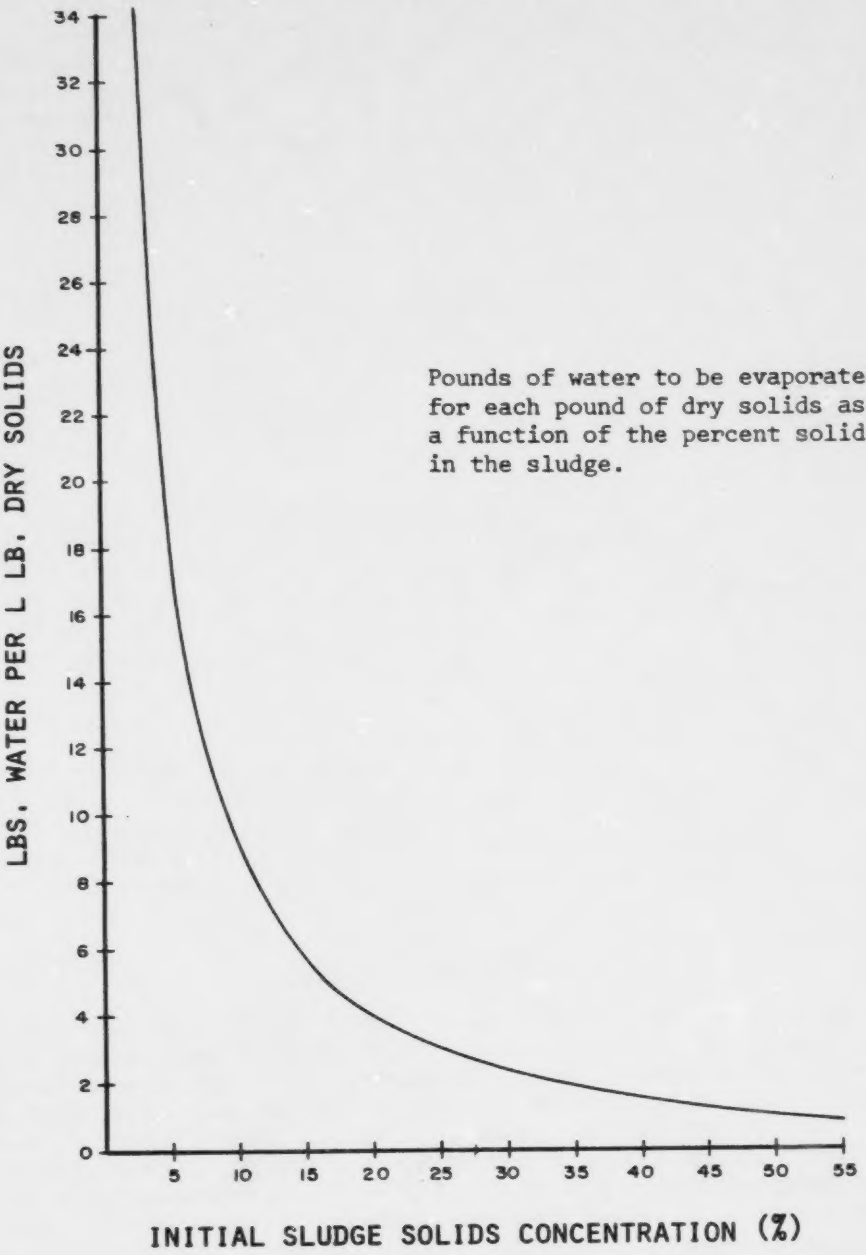
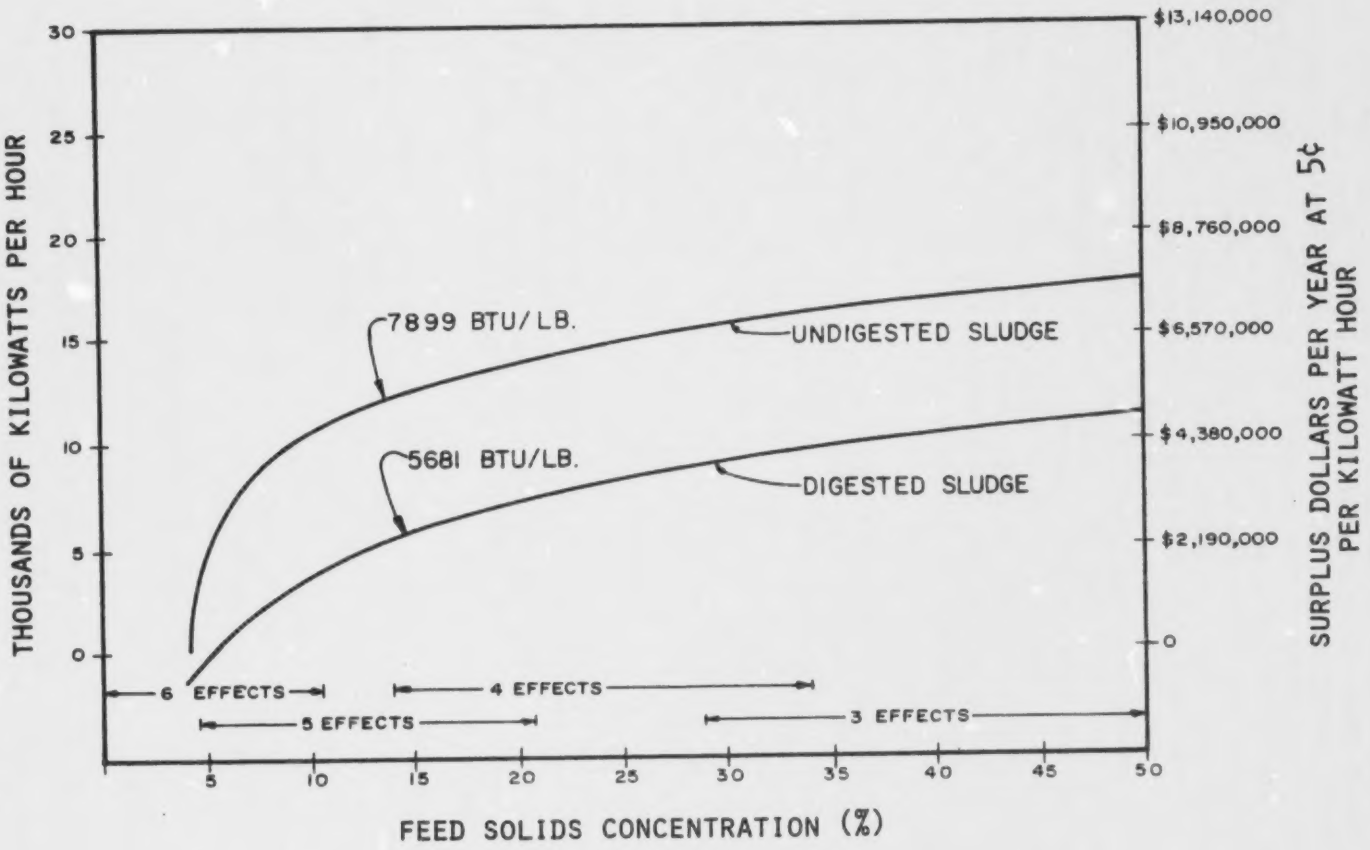


**FIGURE 1.2**

WATER CONTENT VS. PERCENT SOLIDS



Surplus energy recovered as a function of the dry solids concentration. Based on 500 dry tons per day rate. Data supplied by manufacturer.

**FIGURE 1.3**

THE CARVER-GREENFIELD PROCESS
SURPLUS ENERGY RECOVERY

if used for primary sludge, oil losses from the system, quality of the process sidestream (condensate) and cost of operation. It has been mentioned that the City of Omaha had to shut down a Carver-Greenfield installation after operating for a very short period for severe problems. The potential quality of the process sidestream has also been debated in recent years. In a report to the New York--New Jersey Interstate Sanitation Commission, Camp, Dresser and McKee Inc. (1), assumed that the quality of such liquid stream (condensate) is similar in nature to that expected from conventional heat treatment of sludge. They assumed that the cost of sidestream treatment would be $29/ton of dry sludge. However, the manufacturer claimed a much lower concentration of organics and other objectionable constituents in the condensate. Low condensate CODs ( 450 mg/l) were also obtained by Villiers, et al (3) in their exploratory studies employing primary sludge. The cost of the total process has been variously quoted from $13 to $190 per dry ton of sludge processed (6.1), nearly an order of magnitude range.

In 1975, the City of Los Angeles rented a trailer mounted pilot plant from Dehydrotech Corporation and tested it at their Hyperion Treatment Plant (5). The unit was a single-effect evaporator operated as a batch process. Sludges tested were primary sludge both raw and digested, and a blend of raw primary and raw waste-activated sludges. The final product contained 30 to 40 percent process oil and 2 to 5 percent water by weight. However, it was concluded that the bulk of the oil could be recovered from the solids by hydroextraction, a steam distillation technique, which was not a part of the pilot plant set up. While valuable information was derived from this study, there was a need for further information about system performance under a continuous flow mode using both liquid and dewatered sludge as feed material. Data was also lacking about the performance of auxiliary equipment such as the hydroextraction unit.

In light of the above, a continuous flow pilot-scale demonstration study of the Carver-Greenfield process was defined in the LA/OMA Project work program and approved by the LA/OMA Project Policy Board on August 15, 1976. Under a contract from the LA/OMA Project, the City of Los Angeles carried out the demonstration study at their Hyperion Treatment Plant. Results of the demonstration work are described in Chapter 2.

## RESULTS OF OTHER WORK

### CITY OF LOS ANGELES

In 1975, the City of Los Angeles operated a pilot scale Carver-Greenfield unit at their Hyperion Treatment Plant. The types of sludges processed were raw and digested primary sluges, and a blend of undigested primary and waste activated sludges. The following observations were reported by the City (5).

- The process was capable of drying the type of sludges tested to over 95 percent solids (on moisture content basis).

- The dried solids contained 30 to 40 percent oil.

- The process condensate (liquid sidestream) had the following constituent concentrations: oil, 2 mg/l; TDS, 5-16 mg/l; COD, 340 to 1060 mg/l; $NH_3$, 150 to 7170 mg/l. The report concluded that ammonia and COD concentration would actually be lower than these levels in a four-effect system. This is due to the fact that in a four-effect system, evaporation takes place at a relatively lower temperature in the first three effects. The report also concluded that the higher temperatures occuring in a single-effect system would break down more protein, thus releasing more ammonia.

- The fluidizing oil was progressively contaminated, picking up solids from the sludge with each pass through the system.

- Some clogging of the heat exchanger tubes was noticed. Grinding of primary sludge was felt to be essential to prevent such clogging. However, no scaling of the tubes was observed.

- A recommendation was made to test a pilot plant operated in a continuous mode. It was also concluded that the pilot plant should include a sludge grinder and a hydroextraction unit.

## WEYERHAEUSER COMPANY

In 1977, the Weyerhaeuser Company conducted a pilot Carver-Greenfield demonstration study at their Cosmopolis, Washington facility (6). This was done as a part of a series of pilot tests for dewatering their "bio-pond" sludge. The sludge was generated from paper pulp processing activities and was not similar to municipal sludge in characteristics. Therefore, some caution should be exercised in directly applying their findings to municipal sludge handling operation. The following conclusions were drawn from the study:

- A dry granular sludge with solids concentration varying from 42-81 percent was obtained.

- Additional processing was recommended to reduce the residual oil. Steam stripping for volatile oils (petroleum base) and pressing for non-volatile (vegetable) oils were thought to be the most efficient way of achieving this.

- Formation of a gummy phase was observed. The first occurred at a solids content of about 30 percent and another, at about 90 percent T.S. (based on moisture content). However, the report concluded that the problem could be avoided by using proper design and operation considerations.
- Formation of a stable oil-condensate emulsion was noticed with the use of Isopar* as carrying medium. This was traced down to the heavy use of surfactants in the main plant which ended up in the waste stream.
- An odor problem was presented by the non-condensable vapor vented by the vacuum pump.
- The projected total annual cost for sludge drying was $190/ton of dry sludge. The major cost items in the projection were cost of steam for evaporation, residual oil cost, capital amortization, and operation and maintenance.

* An Exxon product.

## ADOLPH COORS COMPANY

While investigating different potential dewatering methods for waste activated sludge, Adolf Coors Company of Golden, Colorado, conducted a pilot study of the Carver-Greenfield process (7).

The Carver-Greenfield process was selected as the most optimum process to deal with their waste activated sludge disposal problem. Recommendations for full-scale construction included use of vegetable oil as a carrying fluid; pressing for extraction of oil from dried sludge and sale of final product as animal feed. A description of their full-scale plant operation has been included in this report in Chapter 3.

# CHAPTER 2

# CHAPTER 2

## REPORT ON CARVER-GREENFIELD PROCESS PILOT DEMONSTRATION

As mentioned in Chapter 1, the LA/OMA Project had decided to conduct a pilot plant investigation of the Carver-Greenfield (C-G) process. A trailer mounted single effect evaporator with other ancillary equipment was rented from Dehydrotech Corporation of East Hanover, New Jersey, the manufacturer of the process. Under a contract from the LA/OMA Project, the City of Los Angeles installed and operated the pilot plant at their Hyperion Treatment Plant. Upon completion of the demonstration work, a report describing the investigation and its results was submitted by the City to the LA/OMA Project (8). The present chapter describes the pilot plant demonstration work and has been based on the same report.

### STUDY OBJECTIVES

Overall objective of this study was to demonstrate on a continuous flow, pilot-scale basis the practicality of the concept of multiple-effect evaporation-pyrolysis of sewage sludges. Sludges which were to be examined during the study included: liquid digested primary sludge, dewatered WAS, dewatered primary and dewatered digested primary. Consideration was to be given to testing sludges from other treatment plants. In all cases sludge was to be processed through Carver-Greenfield evaporation system on a continuous flow single-effect basis, and processed through the hydroextraction unit on a batch basis. Pyrolysis of the final product solids was to be investigated. Specific objectives included:

1. to identify operational problems encountered in the continuous flow system;
2. to develop design and operating parameters for a full-scale system;
3. to obtain capital and operating cost data for a full-scale system;
4. to investigate the efficiency of the hydroextraction process for removal of residual oil in the dewatered solids;
5. to determine the extent of heat exchanger tube fouling during continuous operation and the ability of sludge grinding to mitigate the problem;
6. to identify corrosion problems;
7. to determine the characteristics of final products resulting from the system including liquid sidestreams (i.e., condensate) and exhaust gases;
8. to develop mass balances for the system of Carver-Greenfield, hydroextraction and pyrolysis including fluxes of solids, water and oil;
9. to develop energy balances for the one-effect system including hydroextraction and pyrolysis and extrapolation of these balances to a multiple-effect system.

## METHODS AND MATERIALS

### TEST SET UP

The mobile Carver-Greenfield pilot plant used in this study was rented from Dehydrotech Corporation. Tests were conducted at the City of Los Angeles' Hyperion Treatment Plant, by City forces, with the technical assistance from Dehydrotech, Corporation.

The single-effect pilot unit was originally conceived by Dehydrotech as capable of simulating the multiple-effect feature of the process. By a combination of holding tanks and process piping for flow routing, and by variations in the system pressure (vacuum) and temperature, the single evaporator was to be employed for simulation of two, three or any number of effects, depending on the user's requirements. Initial attempts at this simulation were unsuccessful, including the City of Los Angeles' 1975 testing. Numerous and uncontrolled vacuum leaks and the absence of heat insulation on critical components made simulation difficult. The heat efficiency of the process therefore was not demonstrated by the pilot unit. Some approximations of heat efficiency were explored in the testing.

The pilot plant was moved on-site at Hyperion on May 3, 1977, in the vicinity of the sludge processing facilities, strategically located for sludge feed and the necessary utility connections. Utilities included electricity at 110V and 460V, potable water for cooling, and drainage for system wastes. Process steam was provided by a separately rented boiler. The pilot unit and its ancillary equipment were assembled and all service connections completed on June 1, 1977.

Figure 2.1 is a plot plan of all the test project equipment, including the trailer, the boiler and fuel oil tank, the sludge feed equipment, and the electrical power source. The trailer was delivered to Hyperion with all the equipment mounted in an in-transit condition, ready for assembly. Because the process depends on a tall vertical barometric leg through the heat exchanger to the condenser, a steel framework was supplied with the pilot plant, mounted on the trailer in sections.

The process equipment supplied by Dehydrotech with the trailer unit included: sludge grinder, fluidizing tank with agitator, feed tanks with agitators, heat exchanger, vapor chamber, surface condenser, condensate receiver tanks, recycle oil tank, product tank, vacuum pump, centrifuge, hydroextraction unit, and process pumps. Most of the process instrumentation was housed in a control room at one end of the trailer.

Figure 2.2 describes the are flow schematics of the three operating configurations used for the three types of sludges tested. They all differ in some respect. For the liquid digested and dewatered digested primary sludges, a grinder was included ahead of the feed tanks. The grinder was not needed for waste activated sludge due to its lower content of fibrous and stringy material. The process condensate receiver tanks were redesigned for the

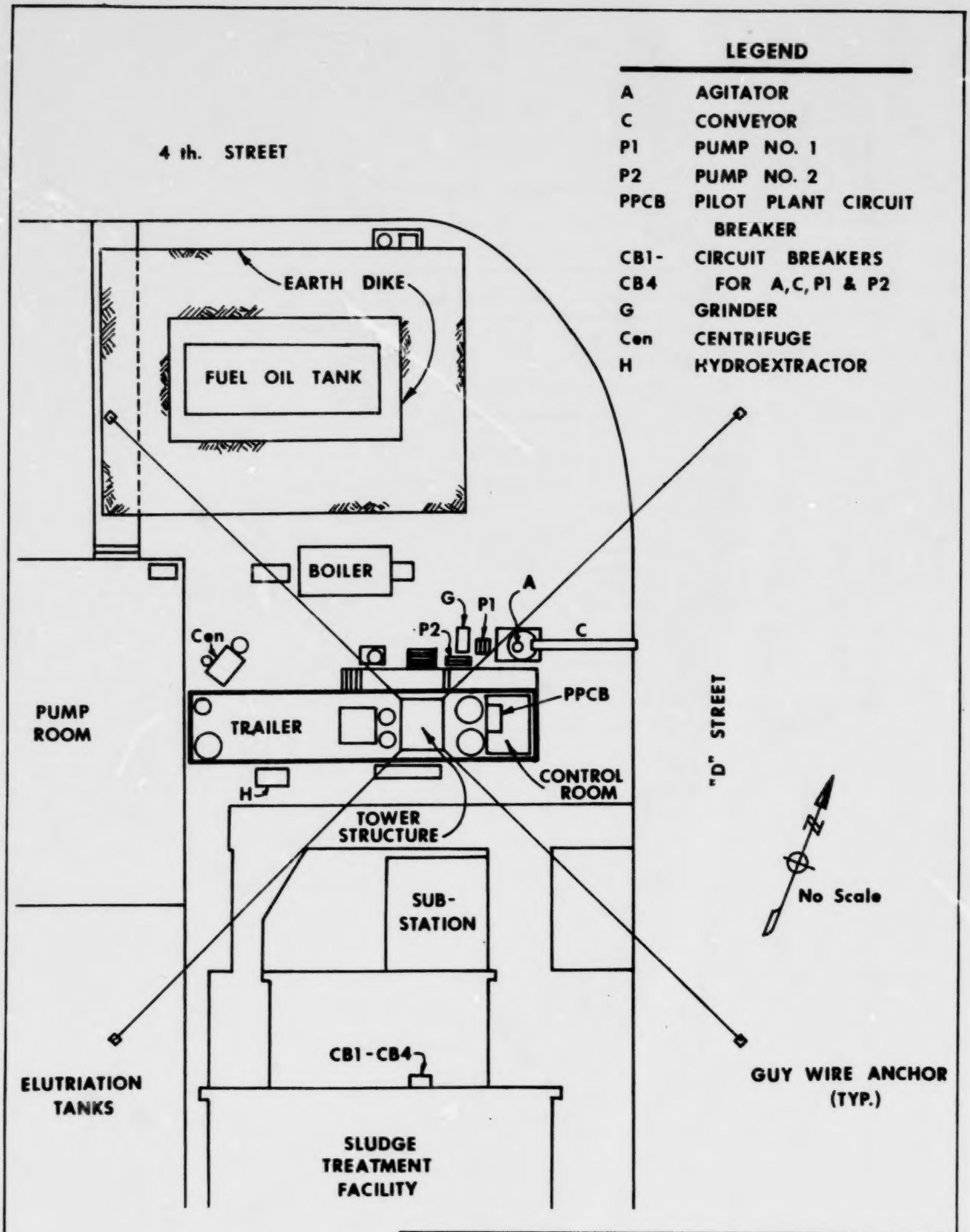
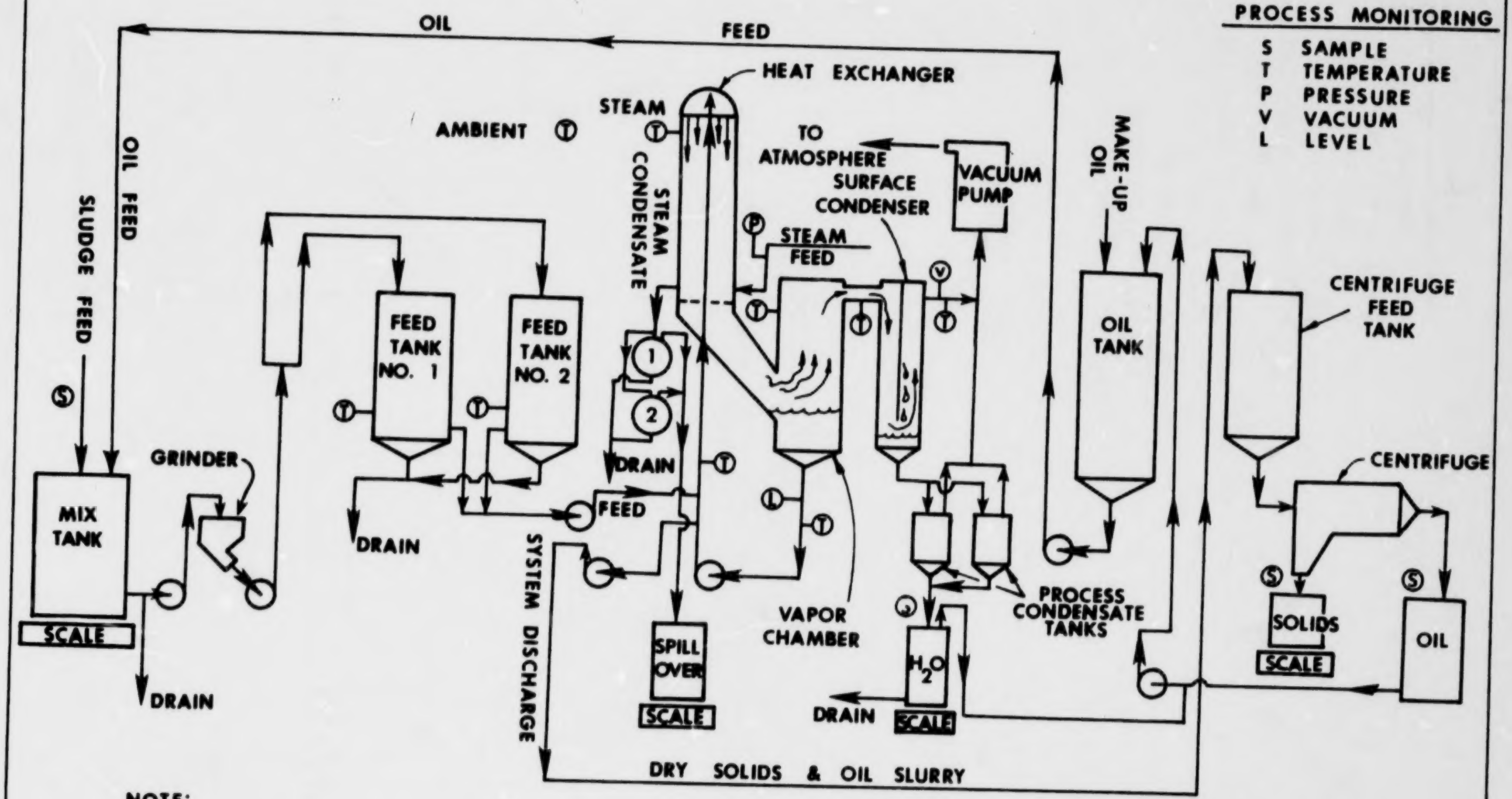

LEGEND
A AGITATOR
C CONVEYOR
P1 PUMP NO. 1
P2 PUMP NO. 2
PPCB PILOT PLANT CIRCUIT BREAKER
CB1- CB4 CIRCUIT BREAKERS FOR A, C, P1 & P2
G GRINDER
Cen CENTRIFUGE
H HYDROEXTRACTOR
4 th. STREET
EARTH DIKE
FUEL OIL TANK
BOILER
G
P1
A
P2
C
Cen
PUMP ROOM
TRAILER
PPCB
CONTROL ROOM
"D" STREET
H
TOWER STRUCTURE
N
No Scale
SUB-STATION
CB1-CB4
ELUTRIATION TANKS
GUY WIRE ANCHOR (TYP.)
SLUDGE TREATMENT FACILITY

FIGURE 2.1

PLOT PLAN FOR CARVER-GREENFIELD PILOT FACILITIES


OIL
FEED
HEAT EXCHANGER
STEAM
AMBIENT T
OIL FEED
SLUDGE FEED
S
MIX TANK
SCALE
DRAIN
GRINDER
FEED TANK NO. 1
FEED TANK NO. 2
T
DRAIN
STEAM CONDENSATE
1
2
DRAIN
FEED
SYSTEM DISCHARGE
SPILL OVER
SCALE
TO ATMOSPHERE
SURFACE CONDENSER
VACUUM PUMP
P
STEAM FEED
V
L
VAPOR CHAMBER
PROCESS CONDENSATE TANKS
$H_2O$
SCALE
DRAIN
DRY SOLIDS & OIL SLURRY
MAKE-UP OIL
OIL TANK
CENTRIFUGE FEED TANK
CENTRIFUGE
SOLIDS
SCALE
OIL
PROCESS MONITORING
S SAMPLE
T TEMPERATURE
P PRESSURE
V VACUUM
L LEVEL

NOTE:
BOILER AND HYDROEXTRACTOR NOT SHOWN.

FIGURE 2.2

CARVER-GREENFIELD PILOT DEMONSTRATION FLOW SCHEMATICS

LA/OMA PROJECT

dewatered digested primary sludge runs, as an aid to continuous operation. The dewatered digested primary set-up required a solids recirculation line (addback) to the mix tank in order to keep the solids in the mix tank in suspension. The mix tank withdrawal point was altered for the waste activated runs.

For the dewatered digested primary schematic the flow courses were as follows. Sludge feed was combined with fluidizing oil in the mix tank, where it was mechanically agitated. The sludge-oil mixture was then pumped to the grinder to macerate any fibrous material which could clog the heat exchanger tube nest downstream. It was then pumped into one of the two feed tanks, where the mixture was also agitated mechanically. Two feed tanks were required for long-term continuous flow operation. Sludge/oil mixture was then pumped through the tube in shell heat exchanger (22 tubes) and was heated by the steam which occupied the heat exchanger jacket. When the heated mixture reached the vapor chamber, water was flash distilled because of its high, temperature and by the action of vacuum applied to the system. Water and oil vapors were recovered from the vacuum chamber and condensed. Water and oil fractions were then gravity separated, the water portion becoming the process condensate. The oil portion represented the more volatile organics in the system, mostly Isopar and were recycled.

The dried sludge, or solids-oil mixture, was withdrawn from the bottom of the vapor chamber. Some of this mixture was recirculated to the heat exchanger to maintain sludge fluidity (addback). The remainder was pumped to the centrifuging operation, where most of the oil was removed from the solids. This recovered oil was pumped back into the oil feed system.

The evaporative capacity of the pilot plant's single-effect system was rated between 200 to 500 pounds of water per hour. Process steam requirements were stated to be around 1000 pounds per hour at 100 psig. The rented boiler, fired by No. 2 fuel oil, had a rated capacity of 1500 to 2500 pounds per hour. The fuel oil storage tank had a capacity of 8100 gallons. The surface condenser required 60 gpm of cooling water at 40 psig. Agitators and the feed and discharge pumps were variable-speed, powered by a hydraulic drive system. The hydroextraction unit was the only batch operated process.

Drums of Isopar L, the fluidizing oil recommended by Dehydrotech for this testing, were purchased from a local supplier. Make-up Isopar was pumped from the drums into the system oil tank, which also received much of the recirculated system oil.

Upon delivery to Hyperion, the pilot plant was not well equipped for continuous mode operation. Modifications were made by Dehydrotech during the first month of testing to correct the deficiencies. Two noteworthy examples were the revisions made to the process condensate withdrawal system, which can be seen from Figure 2.2 and the process instrumentation to allow for continuous monitoring. A continuous operating mode was central to the purpose of this study. The 1975 Hyperion study (5) succeeded in drying a given charge of sludge by application of the necessary heat energy. By comparison, one objective in this study was to draw off dried sludge solids continuously while wet feed was continuously fed to the system.

## TEST PROCEDURES

The planned phases of testing, shown in Table 2.1, called for five and possibly six sludge types. Equipment and time limitations made this goal unattainable. Testing was performed on only three types of sludges. Liquid digested primary sludge was piped to the mix tank from on-site anaerobic digesters. Dewatered digested primary sludge was trucked from the Los Angeles County Sanitation Districts (LACSD) Joint Water Pollution Control Plant in Carson. Thickened waste activated sludge was trucked from the LACSD District 32 Water Renovation Plant in Saugus. The latter two sludge types were not available at Hyperion.

The primary objective of testing was to evaluate the system performance on continuous runs of up to 72 hours. The general schedule of operation was to begin a run on Monday morning, and operate continuously until Thursday, approximately a 72-hour period. Two attempts were made to operate continuously for 10 days. However, both runs were interrupted by equipment malfunctions. Two field technicians from Dehydrotech assisted three engineers from the City of Los Angeles in operating the pilot unit on a round-the-clock three shift basis. At least one engineer from the City and one Dehydrotech technician were on-site during each shift while tests were run.

Test data were recorded on two forms. An operating data sheet was maintained to record the vital system operating parameters including system temperatures and pressures. Sludge feed and process condensate withdrawal were also recorded on the operating data sheet.

A mass balance data sheet was maintained during all test runs to record the weights of all system inputs and outputs. Material fed into the system would include sludge feed, make-up fluidizing oil, initial dry charge material for start-up, etc. Material discharged from the system would include: process condensate, centrifuged solids, centrifuged oil, various spills, estimated evaporation losses from open tanks, etc. These measured weight fluxes were combined with laboratory sample analysis to calculate the mass flow of solids, water and oil entering and leaving the system.

## SAMPLING AND ANALYSIS

Sufficient sampling and analyses were performed to characterize the infeed and outfeed streams from the process, determine process feasibility and potential operating problems, and to determine mass and energy balances to the extent possible.

Regular samples were taken of sludge feed, centrifuged solids, centrifuged oil, and process condensate. Generally, these samples were taken about every two hours. Centrifuged oil and centrifuged solids samples were taken while the centrifuge was operating,

## TABLE 2.1

## PLANNED PHASES OF TESTING

| Phase | Type of Sludge | Duration of Test | Objective | Status |
|---|---|---|---|---|
| 1 | Liquid Digested Primary Sludge | One Week 72-Hour Continuous Run | To determine system performance on a liquid sludge. To estimate system performance for treatment of centrates. | Complete |
| 2 | Dewatered WAS | " | To determine the system capability for dehydration of WAS and the effect of WAS on heat exchanger fouling. To evaluate fertilizer potential of the dry product. | Not run due to inability to obtain sludge. |
| 3 | Dewatered Digested WAS | " | To determine system performance characteristics and the effect of digested WAS on heat exchanger fouling. To evaluate fertilizer potential of the dry product. | Runs completed on DAF thickened digested WAS. |
| 4 | Dewatered Primary | " | To determine the effect of raw primary sludge on heat exchanger fouling. To determine the effect of sludge grinding to reduce heat exchanger fouling. To determine odor problems with processing of raw sludge. | Completed |
| 5 | Dewatered Digested Primary | 2 weeks 240-Hour Continuous Run | To determine the ability of the system to operate on a long-term, continuous flow basis. | Attempted but total operating time less than stated because of equipment problems. |
| 6 | To be Determined | To be Determined | To evaluate other sludges or test conditions as determined by the project team and LA/OMA staff considering information gathered during Phases 1-5. | Second attempt made for 240-hr run. Again limited because of equipment problems. |

normally twice each shift. Sludge feed samples were taken as sludge was added to the mix tank. Process condensate samples were taken from the condensate receiver tank drain, which emptied continuously into a holding drum.

Hydroextraction was performed only once for each type of sludge and was conducted on a batch basis. Samples were taken of the hydroextractor feed (centrifuged solids), hydroextracted solids, hydroextracted oil, and condensed hydroextracted distillate.

All process samples were collected in plastic quart bottles and analyzed in the Hyperion laboratory. Emphasis was placed on obtaining samples representative of full scale operation. Grease, oil and scum layers, and settled solids on tank bottoms were avoided.

A limited number of samples were collected and analyzed by the Los Angeles County Sanitation Districts to document the performance of the process in destroying pathogenic organisms. Pathogen analysis included: total coliform, fecal coliform, fecal streptococus, salmonella, virus and parasitic ova.

Monitoring and sampling of air emissions was scheduled on at least five occasions. Unfortunately, pilot plant malfunction caused postponement on every scheduled date. No air emission data were obtained. However, visual inspection did reveal several potential sources of hydrocarbon emissions which would have to be controlled in full scale operation.

Laboratory analyses were performed on composited samples of the sludge feed, process condensate, centrifuged solids, and centrifuge oil. Except for the analysis of heavy metals and total identifiable chlorinated hydrocarbons (TICH) in the sludge feed and process condensate, composited samples were derived from the samples taken on a 24-hour basis. Roughly, an average of six samples were taken on each process stream each day. At times this increased to as many as nine, and as little as two. Due to the lengthy analysis required for heavy metals and TICH, these parameters were analyzed on a sample composited from all the samples taken for each particular phase of testing. Tables 3 through 10 show the parameters analyzed for the various system inputs and outputs.

Laboratory analysis was performed by Hyperion staff according to Standard Methods(9). The only exception to this was the determination of Isopar content in samples which was conducted using a technique recommended by Dehydrotech (Appendix 1). This technique was necessitated by the high volatility of Isopar. Measured values also included other volatile oils present in the sample.

## RESULTS AND DISCUSSION

### SUMMARY OF PILOT OPERATION

Completed test runs were of generally shorter duration than planned. With every succeeding test run, in fact, the maximum run duration obtainable became shorter, as Table 2.2 shows. A three-day run was obtained for only the liquid digested primary sludge. The longest run with thickened waste activated sludge was 29 hours in spite of efforts to repair malfunctioning equipment. Note that the phase numbering system was modified during actual testing because of the many aborted runs and the disruption of planned testing. Phase designations shown in Table 2.2 will be used in reporting the data throughout this report.

System operating parameters were continuously monitored. When the system was in equilibrium, typical values for all three types of sludges tested were: steam temperature, 250°F; steam pressure, 30 to 40 psig; system vacuum, 13 to 15 inches of mercury; feed tank temperature, 95°F; solid oil slurry recirculation temperature, 240°F; vapor chamber temperature, 210°F; condenser temperature, 210°F entering, and 160°F exiting. The system evaporative rate was generally about 500 pounds per hour, as measured by the process condensate. Ambient temperature for most of the testing was between 75°F and 85°F.

### CONDENSATE QUALITY

As shown in Table 2.3, the range of total dissolved solids was between 16 and 220 mg/l. TDS in the condensate was expected to be low because of the distillation. These values, however, were higher than in the 1975 City of Los Angeles (CLA) study (5) which ranged between 0 and 46 mg/l. This may be due to the fact that samples were taken from this pilot operation at 225°F continuously, rather than at 180°F as in batch drying. Dissolved organics, may be responsible for the TDS increase.

Condensate COD ranged between 212 and 4167 mg/l, higher than in the 1975 CLA study where COD ranged from 103 to 2500 mg/l. As shown in Table 2.3, the higher sludge feed concentrations produced proportionately higher organic concentrations in the process condensate. Similar trends were observed in the 1975 CLA study. However, it should be realized that the moisture content of the feed sludges were vastly different.

When condensate CODs are expressed in terms of pounds of COD per pound of dry sludge feed, as shown in Table 2.4, their differences due to different forms of primary sludge tend to be minimal. The data also indicates that waste activated sludge generates a stronger condensate.

## TABLE 2.2

## SUMMARY OF PILOT PLANT OPERATION

| Phase [a] | Sludge Type | Completed Test Run<br>Inclusive Dates | Completed Test Run<br>Hours of Operation | Remarks |
|---|---|---|---|---|
| 1 | Liquid Digested Primary | 6/6-6/12 | | Testing not continuous, interrupted four times due to equipment malfunction. |
| 1 | " " " | 6/13-6/16 | 76½ | Pilot plant deemed "fully operational" before run. Test run completed. |
| 1A | " " " | 6/20-6/23 | 62½ | Modifications made before run to simplify continuous process monitoring. Test run completed. |
| 2 | Dewatered Digested Primary | 6/27-6/30 | 60 | Test run completed. |
| 3 | " " " | 7/12-7/30 | 57½ | Attempt at 240-hour run. Aborted on 7/12 and 7/14. Extensive modification and repair to hydraulic drive system prior to restarting on 7/28. |
| 3 | " " " | 8/2-8/4 | 46½ | Second attempt at 240-hour continuous run. Equipment failure again caused reduced run. |
| 4 | Thickened Waste Activated | 8/16 | 11 | Aborted run on 8/11 due to equipment failure. |
| 4 | " " " | 8/18-8/19 | 29 | Failed to run for 72-continuous hours due to equipment failures. |
| 4 | " " " | 8/22 | 4½ | Last attempt for 72-hour run. Mechanical failures not repairable. |

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[a] These phase designations do not agree with the planned phases of operation as described in Table 1. The phase designations in this table are the ones used in reporting the data for this study.

## TABLE 2.3

## PROCESS CONDENSATE CHEMICAL ANALYSIS

| Duration of Run | Phase | No. of Samples in Comp. | TDS mg/1 | Volatile Oil % | $NH_3$-N mg/1 | pH | $BOD_5$[a] mg/1 | COD[a] mg/1 | Vol. Acids ppm |
|---|---|---|---|---|---|---|---|---|---|
| **Liquid Digested Primary** | | | | | | | | | |
| 6/13-6/14 | 1 | 16 | 60 | 0.2 | 890 | 9.5 | 207 | 521 | 245 |
| 6/14-6/15 | 1 | 12 | 80 | 1.0 | 911 | 9.4 | 150 | 234 | 98 |
| 6/15-6/16 | 1 | 12 | 32 | 0.8 | 822 | 9.4 | 222 | 396 | 270 |
| Weighted Average | | | 58 | 0.6 | 876 | 9.4 | 194 | 397 | 208 |
| 6/20-6/21 | 1A | 7 | 54 | 0.6 | 817 | 8.6 | 294 | 506 | 262 |
| 6/21-6/22 | 1A | 11 | 52 | 0.2 | 783 | 9.3 | 144 | 212 | 93 |
| 6/22-6/23 | 1A | 11 | 66 | 0.5 | 708 | 9.2 | 198 | 383 | 234 |
| Weighted Average | | | 58 | 0.4 | 763 | 9.1 | 201 | 348 | 187 |
| **Dewatered Digested Primary** | | | | | | | | | |
| 6/27-6/28 | 2 | 4 | 46 | 3.4[b] | 775 | 9.4 | 1440 | 1420 | 609 |
| 6/28-6/29 | 2 | 8 | 220 | 1.1 | 1270 | 9.5 | 2460 | 2271 | 227 |
| 6/29-6/30 | 2 | 9 | 32 | 0.2 | 1498 | 9.3 | 3240 | 3400 | 1616 |
| Weighted Average | | | 106 | 0.6 | 1273 | 9.4 | 2600 | 2593 | 895 |
| 7/28-7/29 | 3 | 4 | 150 | 0.0 | 1749 | 9.6 | 5250 | 4167 | 2779 |
| 7/29-7/30 | 3 | 7 | 20 | 0.0 | 1592 | 9.7 | 4354 | 2886 | 1450 |
| 7/30 | 3 | 4 | 50 | 0.0 | 1434 | 9.6 | 4354 | 1856 | 973 |
| 8/2-8/3 | 3 | 9 | 34 | 0.0 | 1707 | 9.6 | 2610 | 1624 | 876 |
| 8/3-8/4 | 3 | 12 | 16 | - | 1585 | 9.7 | 1740 | 914 | 485 |
| Weighted Average | | | 40 | 0.0 | 1618 | 9.7 | 3146 | 1941 | 1080 |
| **Thickened Waste Activated** | | | | | | | | | |
| 8/18 | 4 | 4 | - | 0.0 | 741 | - | - | - | 2714 |
| 8/19 | 4 | 9 | 46 | 0.2 | 637 | 8.6 | 2250 | 2240 | 1948 |
| 8/22 | 4 | 2 | 50 | 0.0 | 464 | 8.7 | 959 | 1340 | 832 |
| Weighted Average | | | 47 | 0.1 | 642 | 8.6 | 2015 | 2076 | 2003 |

a/ Ratio of $BOD_5$ to COD for five phases (weighted averages) were: 0.5, 0.6, 1.0, 1.6 and 1.0.

b/ Not included in weighted average. Considered invalid.

## TABLE 2.4

## CONDENSATE CHEMICAL OXYGEN DEMANDS

| FEED | CHEMICAL OXYGEN DEMAND mg/l | CHEMICAL OXYGEN DEMAND Lb/Lb of dry sludge |
|---|---|---|
| Liquid Digested Primary | 350 | 0.012 |
| Dewatered Digested Primary | 2600 | 0.008 |
| Thickened Waste Activated | 2100 | 0.08 |

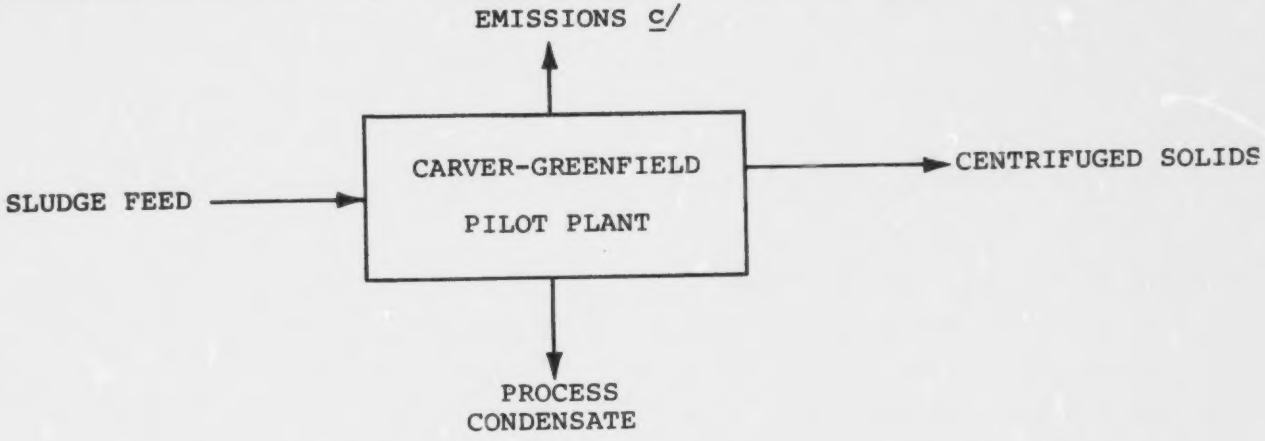
The range of $BOD_5$ was 144 to 5250 mg/l, as shown in Table 2.3. The ratio of COD to BOD values indicates that most of the COD is biodegradable, an important consideration in the recirculation of the process condensate to the treatment plant. The measured condensate BOD for the dewatered digested primary was higher than the COD. It is conceivable that the $BOD_5$ measurements were influenced by the high ammonia content in the samples. However, the anomaly cannot be fully explained.

Visually, the condensate was a clear liquid, without any noticeable color cast. The condensate has a definite ammonia odor. It was easily separated from Isopar which might have distilled with it, due to the immiscibility of the two liquids.

Nitrogen as ammonia ranged from 464 to 1749 mg/l. The 1975 CLA results had a wider range: 150 to 7170 mg/l. Based on the ammonia levels in the sludge feed (Table 2.6) 80 to 95 percent of the feed ammonia distilled over into the process condensate, as shown in Figure 2.3.

Heavy metal concentrations in the process condensate are shown in Table 2.5. Heavy metal concentrations are generally very low. Nine of the ten metals tested had concentrations of 0.1 mg/l or less.

Volatile oil content of the process condensate (which included both Isopar and lighter fractions of sludge oils) ranged between a non-detectable amount to 1.1 percent as shown in Table 2.3. Higher concentrations occurred in Phases 1, 1A and 2, before the condensate withdrawal system was modified for continuous flow operation. Oil contamination was a concern in taking condensate samples and apparently was not effectively avoided before such modification was made After modification, (Phase 3 to the finish of testing) condensate oil content dropped dramatically to expected levels. It should be noted that an oil content in the process condensate of about 1 percent would be excessive, and would call for the inclusion of a coalescing unit to separate and recover these remaining oils. A coalescing unit was not included in the pilot system and the oil content shown in Table 2.3 are the result of gravity separation only It would appear that a coalescing unit would be required to assure minimal loss of fluidizing oil in the process condensate.


EMISSIONS c/
SLUDGE FEED
CARVER-GREENFIELD PILOT PLANT
CENTRIFUGED SOLIDS
PROCESS CONDENSATE

| | Sludge Feed | | | | | Process Cond. $NH_3$-N | | | Centr. Solids | | $NH_3$-N |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | $NH_3$-N | | | | | | | |
| Sludge Type | T.S. a/ % | Solid b/ lb. | $H_2O$ b/ lb. | ppm. | lb. | Conc. mg/l | Wt. d/ lb. | % of Feed | Conc. mg/l | Wt. d/ lb. | % of Feed |
| Liquid Digested Primary | 2.76 | 0.028 | 0.972 | 917 | 0.00092 | 763 | 0.00074 | 80 | 94 | 0.0000026 | 0.3 |
| Dewatered Digested Primary | 24.10 | 0.241 | 0.759 | 1024 | 0.00102 | 1273 | 0.00097 | 95 | 324 | 0.0000780 | 7.6 |

a/ Weighted average of selected run.

b/ Assuming one pound of wet feed sludge.

c/ $NH_3$-N not measured in pilot study.

d/ Weight x concentration x $10^{-6}$.

**FIGURE 2.3**

AMMONIA MASS BALANCE ANALYSIS

## TABLE 2.5

## PROCESS CONDENSATE CHEMICAL ANALYSIS

(mg/l)

| Constituent | Liquid Dig. Primary | | Dewatered Dig. Primary | | Thick. WAS |
|---|---|---|---|---|---|
| | Phase I<br>6/13-6/16 | Phase IA<br>6/20-6/23 | Phase 2<br>6/27-6/30 | Phase 3<br>7/28-8/4 | Phase 4<br>8/16-8/19 |
| Ag | 0.01 | 0.01 | 0.01 | 0.00 | 0.00 |
| As | 0.00 | 0.00 | 0.00 | 0.00 | - |
| Cd | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| CN | 0.00 | 0.01 | 0.05 | 0.02 | 0.02 |
| Cr | 0.06 | 0.04 | 0.02 | 0.01 | 0.01 |
| Cu | 0.08 | 0.04 | 0.05 | 0.10 | 0.02 |
| Hg | 0.0027 | 0.0031 | 0.0037 | 0.0019 | - |
| Ni | 0.04 | 0.03 | 0.03 | 0.05 | 0.02 |
| Pb | 0.01 | 0.01 | 0.00 | 0.02 | 0.01 |
| Zn | 0.14 | 0.03 | 0.10 | 0.56 | 0.10 |
| TICH | 0.0020 | 0.0012 | 0.0019 | 0.0013 | 0.0043 |

NOTE: Analyses made on a single sample composited from samples collected daily.

## FEED AND PRODUCT CHARACTERISTICS

Chemical characteristics of the sludge feed are summarized in Tables 2.6 and 2.7. Approximate feed solids concentrations were as follows: liquid digested primary, 2.5 percent; dewatered digested primary, 25 percent; and thickened waste activated (WAS), 2.5 percent. Analysis was made for both "oil and grease" and for "volatile oils". The former is a conventional parameter. The latter was the result of laboratory analysis of the sludge feed using the recommended method of analysis for Isopar. Theoretically there should not be any Isopar present in the sludge feed (although contamination in the pilot system was possible), So the measured values represent highly volatile oils present in the original sludge feed. This volatile oil content is significant in that it is a rough measure of the lighter oils contained in the sludge itself which would partially compensate for Isopar losses through the system.

Measured concentrations of metals and chlorinated hydrocarbons were within the ranges observed from normal plant operating data. If any anomalies exist between the three sludge types, partial explanation could be found in the fact that the liquid digested primary was obtained from the City's Hyperion Plant, and the other sludges were obtained from plants operated by the Los Angeles County Sanitation Districts (LACSD).

A complete table of properties for Isopar L is included in Appendix I. Isopar L is ideally suited as a fluidizing medium for several reasons. It is azeotropic; it has the property of vaporizing at a temperature below its boiling point in the presence of steam. It is extremely immiscible in water, with a solubility of less than one part per billion in water. Isopar is a food grade oil.

Analyses of the centrifuged solids are summarized in Table 2.7 and 2.8. Residual oil ranged between 37.2 and 47.4 percent, total solids between 51.3 and 60.9 percent, and moisture between 1.0 and 7.7 percent. The solids concentration computed on an oil-free basis is indicative of the effectiveness of the process in removing water from the sludge. On an oil-free basis, solids concentration of the centrifuged solids ranged between 87.2 and 98.1 percent, with an average value of 95.7 percent. Centrifuged solids were generally black in color and were odorous. The solids had a uniform consistency and could probably be conveyed with little or no difficulty. The digested primary centrifuged solids could be picked up as a cohesive mass because of the fibrous material contained in primary sludges. The waste activated centrifuged solids were more particulate, tending to fall apart somewhat like sand when handled. Properties of the centrifuged oil are discussed later under oil recovery systems.

In a full scale system, it would be necessary to remove additional oil beyond centrifugation to prevent excessive losses of Isopar with the centrifuged solids. The hydroextraction process was tested as a means for additional oil recovery. Hydroextraction involves distillation of volatile oils by the application of superheated steam.

## TABLE 2.6

## SLUDGE FEED CHEMICAL ANALYSIS

| Duration of Run | Phase | No. of Samples in Comp. | Total Solids % | Volatile Content % | Volatile Oils[a] % | $H_2O$ % | pH | $NH_3$-N[b] mg/l | Org N[b] mg/l | Oil and Grease[b] mg/l |
|---|---|---|---|---|---|---|---|---|---|---|
| Liquid Digested Primary | | | | | | | | | | |
| 6/13-6/14 | 1 | 5 | 2.59 | 66.0 | 3.6 | 91 | 6.9 | 1003 | 1175 | 1860 |
| 6/14-6/15 | 1 | 6 | 2.48 | 64.1 | 4.2 | 90 | 7.3 | 933 | 1019 | 1540 |
| 6/15-6/16 | 1 | 6 | 2.64 | 64.8 | 0.2 | 97 | 9.1 | 928 | 1050 | 1680 |
| Weighted Average | | | 2.57 | 64.9 | 2.6 | 93 | 7.8 | 952 | 1076 | 1684 |
| 6/20-6/21 | 1A | 6 | 2.96 | 62.2 | 0.0 | 93 | 7.4 | 937 | 742 | 1720 |
| 6/21-6/22 | 1A | 5 | 2.62 | 62.6 | 0.0 | 94 | 7.2 | 895 | 1075 | 1600 |
| 6/22-6/23 | 1A | 3 | 2.58 | 63.2 | 0.2 | 93 | 7.1 | 914 | 1072 | 1500 |
| Weighted Average | | | 2.76 | 62.6 | 0.0 | 93 | 7.3 | 917 | 932 | 1630 |
| Dewatered Digested Primary | | | | | | | | | | |
| 6/27-6/28 | 2 | 3 | 23.9 | 53.5 | 0.0 | 75 | - | 926 | 4959 | 9120 |
| 6/28-6/29 | 2 | 7 | 24.2 | 53.1 | 2.3 | 72 | - | 1035 | 4866 | 10350 |
| 6/29-6/30 | 2 | 7 | 24.1 | 51.2 | 0.0 | 75 | - | 1056 | 10000 | 9320 |
| Weighted Average | | | 24.1 | 52.4 | 0.9 | 74 | - | 1024 | 6996 | 9709 |
| 7/28-7/29 | 3 | 6 | 24.2 | 52.3 | 0.0 | 75 | - | 1247 | 4902 | - |
| 7/29-7/30 | 3 | 5 | 24.5 | 51.3 | 0.0 | 75 | - | 1451 | 6130 | - |
| 7/30 | 3 | 2 | 27.5 | 51.9 | 0.0 | 72 | - | 1166 | 6355 | - |
| 8/2-8/3 | 3 | 10 | 28.1 | 51.9 | 0.0 | 71 | - | 1784 | 6634 | - |
| 8/3-8/4 | 3 | 9 | 26.5 | 49.2 | 0.0 | 73 | - | 1681 | - | - |
| Weighted Average | | | 26.3 | 51.1 | 0.0 | 73 | - | 1564 | 6048 | - |
| Thickened Waste Activated | | | | | | | | | | |
| 8/19 | 4 | 2 | 2.55 | 78.8 | 5.0 | 95 | - | 465 | 1329 | - |

[a] See Appendix I for method of analysis.
[b] Expressed in mg/kg for dewatered primary sludges.

## TABLE 2.7

## SLUDGE FEED CHEMICAL ANALYSIS

(mg/l)

| Constituent | Liquid Dig.Primary | | Dewatered Dig.Primary | | Thick.WAS |
|---|---|---|---|---|---|
| | Phase 1<br>6/13-6/16 | Phase 1A<br>6/20-6/23 | Phase 2<br>6/27-6/30 | Phase 3<br>7/28-8/4 | Phase 4<br>8/16-8/19 |
| Ag | 5.36 | 4.72 | 1.40 | 9.72 | 0.59 |
| As | 0.55 | 0.62 | 3.29 | 2.46 | 0.02 |
| Cd | 3.60 | 3.94 | 14.0 | 17.6 | 0.32 |
| Cn | 0.01 | 0.04 | 0.14 | 0.34 | 0.07 |
| Cr | 56 | 56 | 500 | 539 | 8.48 |
| Cu | 45 | 44 | 243 | 272 | 15 |
| Hg | 0.34 | 0.50 | 13.4 | 3.61 | 0.22 |
| Ni | 10.4 | 10.0 | 102 | 90 | 1.49 |
| Pb | 6.54 | 10.5 | 230 | 101 | 2.50 |
| Zn | 91 | 94 | 800 | 730 | 30 |
| TICH | 0.0049 | 0.0077 | 0.196 | 0.227 | 0.273 |

NOTE: Analyses made on a mixture of the daily composited samples.
All units in mg/l.

## TABLE 2.8

## CENTRIFUGED SOLIDS CHEMICAL ANALYSIS

| Duration of Run | Phase | No. of Samples in Comp. | Total Solids % | Volatile Content % | Oil % | $H_2O$ % | $NH_3$-N[b] mg/l | Org N[b] mg/l |
|---|---|---|---|---|---|---|---|---|
| Liquid Digested Primary | | | | | | | | |
| 6/13-6/14 | 1 | 4 | 52.4 | 62.4 | 37.4 | 4.2 | 184 | 4555 |
| 6/14-6/15 | 1 | 10 | 51.5 | 61.4 | 39.8 | 4.2 | 256 | 620 |
| 6/15-6/16 | 1 | 8 | 52.5 | 62.1 | 33.9 | 7.7 | 216 | 119 |
| Weighted Average | | | 52.0 | 61.8 | 37.2 | 5.5 | 228 | 1765 |
| 6/20-6/21 | 1A | 4 | 55.5 | 56.4 | 42.9 | 1.5 | 140 | 37 |
| 6/21-6/22 | 1A | 10 | 51.3 | 58.9 | 47.4 | 1.0 | 98 | 11 |
| 6/22-6/23 | 1A | 9 | 53.7 | 59.9 | 44.0 | 1.5 | 68 | 39 |
| Weighted Average | | | 53.0 | 58.9 | 45.3 | 1.3 | 94 | 29 |
| Dewatered Digested Primary | | | | | | | | |
| 6/27-6/28 | 2 | 4 | 57.9 | 55.6 | 47.0 | 2.0 | 207 | 0 |
| 6/28-6/29 | 2 | 8 | 57.9 | 52.0 | 41.6 | 1.8 | 325 | 20 |
| 6/29-6/30 | 2 | 9 | 57.4 | 48.9 | 40.5 | 2.2 | 374 | 53 |
| Weighted Average | | | 57.7 | 51.4 | 42.2 | 2.0 | 324 | 24 |
| 7/28-7/29 | 3 | 5 | 54.7 | 51.4 | 43.4 | 1.2 | 538 | 40 |
| 7/29-7/30 | 3 | 5 | 58.2 | 50.6 | 40.6 | 1.2 | 1497 | 94 |
| 7/30 | 3 | 3 | 56.6 | 52.0 | 42.9 | 1.5 | 524 | 7 |
| 8/2-8/3 | 3 | 9 | 60.9 | 51.6 | 37.3 | 1.5 | 768 | 3416 |
| 8/3-8/4 | 3 | 12 | 56.2 | 49.2 | 40.6 | 2.5 | 780 | 33 |
| Weighted Average | | | 57.6 | 50.6 | 40.3 | 1.8 | 824 | 718 |
| Thickened Waste Activated | | | | | | | | |
| 8/19 | 4 | 5 | 55.9 | 68.5 | 41.2 | 4.0 | 501 | 7165 |
| 8/22 | 4 | 2 | 56.3 | 69.5 | 38.3 | 10.0[a] | 177 | 13000 |
| Weighted Average | | | 56.0 | 68.8 | 40.4 | 5.7 | 408 | 10083 |

[a] Value not used in analysis. Considered invalid.

[b] Expressed in mg/kg for dewatered primary sludges.

The hydroextraction unit was operated on a batch basis, and results are given in Table 2.9. Residual oil in the hydroextracted solids was about 1 percent or less. Although a continuous flow hydroextraction operation may not necessarily perform as well, the batch testing should verify the potential of the basic process of steam stripping oils from the sludge solids. Total solids concentration of the hydroextracted material ranged between 92.7 and 99.5 percent. Residual moisture ranged between 0.5 and 2.7 percent. Hydroextracted solids had a powdery consistency and could create a dust problem in conveyance and hauling unless properly handled.

## OIL RECOVERY SYSTEMS

Solids content of the centrifuged oil reflects the solids capture efficiency of the dewatering process. Concern has been expressed over the potential build-up of fines in the recirculated oil. As shown in Table 2.10, average solids content ranged between 2.7 and 10.8 percent. The solids content both rose and fell as the testing for each phase progressed, allowing for no definitive conclusion as to the potential build-up of fines under long term continuous operation. The high solids measured in the centrifuged oil was also partly due to the method of sample analysis. The solids test procedure used measured all material which did not volatilize at 103°C, which included heavy oils as well as solids.

The centrifuge oil was black in color and had an offensive odor, compared to the colorless and odorless Isopar. Sewage organics undoubtedly caused the odor. There appeared to be little difference in viscosity between the centrifuged oil and Isopar. It should be noted that the recirculated oil in an operational system would not consist of pure Isopar. The system would attain an equilibrium condition in which a generally fixed proportion of the system oil would be made up of lighter sludge oils.

Hydroextraction was performed on a batch basis, one test for each of Phase IA, 2, 3 and 4. Table 2.9 summarized the laboratory analyses made on the hydroextractor feed (centrifgued solids), hydroextracted solids, and hydroextracted oil. Comparison of the oil content in the centrifuged feed and the hydroextracted solids show that hydroextraction is an effective oil recovery process. The solids content of the hydroextracted oil was approximately between 1 and 5 percent. This is a lower solids content in general than in the centrifuged oil.

Dust contamination in the condensing unit of the hydroextractor was a recurrent problem. The first samples obtained from the condensing unit were clear, but the oil in most of the subsequent samples appeared black. Apparently, the solids became dust after distillation of the oils. Conditions in the hydroextractor then allow some of the dust to escape with the distilled oil vapors. Such fine solids would ultimately appear in the condensed oil which is recycled to the system or in the vent air which exits the condenser. In a commercial operation the hydroextraction system should be designed to avoid unnecessary turbulence to reduce the dusty problem. Nevertheless, it may not be entirely avoidable. Care would have to be exercised to avoid explosive dust concentrations in vent air from the condenser.

## TABLE 2.9
## HYDROEXTRACTION RESULTS

| Date | Phase | Total Solids % | Volatile Content % | Oil % | $H_2O$ % | $NH_3$-N mg/kg | Org N mg/kg |
|---|---|---|---|---|---|---|---|
| Centrifuged Feed | | | | | | | |
| 6/23 | 1A | 55.1 | 59.0 | 40.6 | 1.5 | -- | -- |
| 6/30 | 2 | 58.3 | 51.2 | 39.5 | 2.0 | -- | -- |
| 8/3 | 3 | 56.8 | 49.8 | 38.3 | 1.8 | -- | -- |
| 8/21 | 4 | 60.9 | 73.0 | 34.4 | 2.3 | -- | -- |
| Hydroextracted Solids | | | | | | | |
| 6/23 | 1A | 99.5 | 58.2 | 0.5 | 0.5 | -- | -- |
| 6/30 | 2 | 98.1 | 50.6 | 0.2 | 1.0 | -- | -- |
| 8/4 | 3 | 92.7 | 48.6 | 1.1 | 2.7 | 600 | 36,000 |
| 8/21 | 4 | 99.3 | 65.8 | 0.0 | 0.9 | 380 | 45,000 |
| Hydroextracted Oil | | | | | | | |
| 6/23 | 1A | 4.77 | | 49.7 | 42.0 | | |
| 6/30 | 2 | 3.19 | | 92.4 | 3.2 | | |
| 8/4 | 3 | 0.92 | | - | 0.0 | | |
| 8/21 | 4 | 1.23 | | 86.2 | 1.1 | | |

A liquid sidestream from the hydroextraction process results from the moisture in the centrifuged solids which distills with the lighter oils, and is collected in the condensate. Table 2.10 summarized the laboratory results of the analyses made on this liquid sidestream. Representative samples were difficult to obtain because of the emulsion of dust in the condensed vapors and an indistinct oil/water interface in the condensing unit. Values shown in Table 2.11 reflect this contamination to varying degrees. The quantities of hydroextracted condensate is small in proportion to the condensate produced in the evaporation step. Therefore, it would likely have a negligible effect on the wastewater plant if combined with the process condensate and recycled. Except for TDS concentrations, this condensate is similar to the process condensate in character.

## HEAVY OIL SEPARATION

Recycle oil separated from the dry sludge solids in the centrifuge is high in total solids and heavy oil from the sludge. To prevent build-up of these constituents in the fluidizing oil, a steam stripper would be employed on a full scale plant. In order to test the ability of a stripper to do this, a small quantity of recycle oil was sent through the pilot plant single-effect evaporator under conditions simulating those of a commercial steam stripper. Total solids content of the recycle oil feed was 6.9 percent. Total solids content of the stripped fluidizing oil was 0.02 percent, and that of the separated heavy oil, 21.3 percent. The high solids content in the heavy oil may be misleading because the total solids test according to Standard Methods(9) measures all material which is not volatilized at 103°C. Since the heavy oil residue after heating was observed to be fluid, it was concluded that the total solids test measured the non-volatilized portion of oil as well as the solids. In any event, the measured solids content of the stripped fluidizing oil shows that stripping would aid in preventing the build-up of heavy oils from the sludge.

However, the test data do not indicate an efficient separation of Isopar from the heavy oil. Almost 79 percent of the heavy oil fraction volatilized at 103°C (217°F) during the total solids analysis. For an efficient distillation of Isopar a temperature of around 270°F is required in the steam stripper. Any material that does not volatilize at this temperature is left behind and termed as heavy oil. The heavy oil fraction thus separated should not be expected to give up 79 percent of its weight when heated to 217°F (103°C) (which is considerably lower than 270°F). The reason for such inefficient distillation of Isopar may be attributed to the fact that high pressure steam (150 psig) was not available for this test. However, fractional distillation being a standard industrial unit operation, separation of Isopar from heavy oil can be easily achieved in a properly designed plant.

## HEAT EXCHANGER FOULING

Potential fouling of the heat exchanger may take two forms: clogging of the top entrance to the 1¼-inch diameter tubes by fibrous material; and scaling of the inside walls of the tubes due to the baking

## TABLE 2.10

## CENTRIFUGED OIL CHEMICAL ANALYSIS

| Duration of Run | Phase | No. of Samples in Comp. | Total Solids % | Volatile Content % of T.S. | pH |
|---|---|---|---|---|---|
| Liquid Digested Primary | | | | | |
| 6/13-6/14 | 1 | 6 | 0.04 | 50.0 | 5.5 |
| 6/14-6/15 | 1 | 10 | 3.19 | 88.1 | 5.9 |
| 6/15-6/16 | 1 | 8 | 4.21 | 87.6 | 6.8 |
| Weighted Average | | | 2.74 | 78.4 | 6.1 |
| 6/20-6/21 | 1A | 4 | 5.85 | 86.3 | 6.3 |
| 6/21-6/22 | 1A | 10 | 5.79 | 85.3 | 6.7 |
| 6/22-6/23 | 1A | 10 | 4.99 | 89.0 | 6.6 |
| Weighted Average | | | 5.47 | 87.0 | 6.6 |
| Dewatered Digested Primary | | | | | |
| 6/27-6/28 | 2 | 3 | 4.29 | 83.7 | 6.4 |
| 6/28-6/29 | 2 | 8 | 3.82 | 81.4 | 6.4 |
| 6/29-6/30 | 2 | 9 | 5.56 | 74.8 | 7.1 |
| Weighted Average | | | 4.67 | 78.8 | 6.7 |
| 7/28-7/29 | 3 | 3 | 4.13 | 79.4 | 7.2 |
| 7/29-7/30 | 3 | 5 | 7.3 | 76.3 | 7.8 |
| 7/30 | 3 | 3 | 9.3 | 74.9 | 7.6 |
| 8/2-8/3 | 3 | 9 | 9.76 | 76.9 | 8.3 |
| 8/3-8/4 | 3 | 12 | 13.4 | 80.6 | 8.8 |
| Weighted Average | | | 10.2 | 78.2 | 8.2 |
| Thickened Waste Activated | | | | | |
| 8/19 | 4 | 5 | 12.4 | 74.5 | 6.1 |
| 8/22 | 4 | 2 | 6.7 | 79.4 | 6.7 |
| Weighted Average | | | 10.8 | 75.9 | 6.3 |

## TABLE 2.11

## HYDROEXTRACTION PROCESS CONDENSATE

| Date | Phase | | TDS mg/l | Oil % | $NH_3$-N mg/l | pH | $BOD_5$ mg/l | COD mg/l | Vol. Acids ppm |
|---|---|---|---|---|---|---|---|---|---|
| 6/23 | 1A | black sample a/ | 1486 | 48.1 | 235 | 7.7 | 1195 | b/ | 112 |
| | | clear sample | 136 | 0.2 | 129 | 8.8 | 204 | 360 | 251 |
| 6/30 | 2 | | 524 | 0.2 | 476 | 7.7 | 2640 | 2745 | 1413 |
| 8/4 | 3 | black sample a/ | 254 | - | 525 | 9.3 | b/ | b/ | 794 |
| | | clear sample | 196 | - | 980 | 9.4 | 2130 | 2020 | 790 |
| 8/21 | 4 | | 2660 | 0.0 | 502 | 7.2 | 4000 | 6170 | 2150 |

a/ Analysis was made of the dark and clear strata when the sample was allowed to stand still.

b/ Denotes too much oil in sample for analysis.

of sludge solids. The tubes were flushed out prior to the start of the first run on Phase 1A as part of a final system purge. The heat exchanger was not disassembled or cleaned for the duration of testing, a total run time of about 350 hours. Heat exchanger fouling was not experienced during this run time.

After all testing had been completed, the heat exchanger was carefully disassembled, and no obstructions were found at the tube entrance. Primary sludges are potentially the most troublesome because they contain fibrous material that tends to bridge the 1¼-inch diameter tubes, and form a mat-like barrier. Disintegration of the sludge feed with a sludge grinder eliminated the clogging problem experienced in the 1975 CLA study. Waste activated sludges may not pose a problem in this respect and the grinder was not used for testing of the waste activated sludge in Phases 3 and 4.

Examination of the heat exchanger tube interiors revealed that there was no scaling on the walls. This corroborated results of the 1975 CLA testing (5) and indicated that use of a fluidizing oil does protect the heat exchangers from scaling and corrosion.

## OIL:SOLIDS RATIO

The ratio of fluidizing oil to sludge solids recommended for this testing by Dehydrotech was 10:1. Laboratory analyses of samples taken from the two feed tanks indicated that the ratio was well above 10:1 for Phases 1 and 1A, the liquid digested primary runs. Poor system circulation was experienced in Phase 1, diagnosed as excessive boiling off of Isopar with the sludge water. Apparently, the high oil to solids ratios reflect operator overcompensation in adding large quantities of Isopar to the system to overcome the problem. The situation improved in Phase 1A, and in the remaining phases of operation the ratios decreased to generally below 10:1. It is believed that the high ratios experienced in the liquid digested primary runs were due to the operator's inexperience in maintaining system equilibrium and the observation made during the initial phase was not indicative of process requirement.

In the 1975 CLA study (5), oil to solids ratios of up to 20:1 were used for primary sludge and for mixtures of primary and waste activated sludges. On test runs with waste activated sludge alone, ratios of 12:1 and less were used. Higher ratios were required for the feeds containing primary sludge in order to prevent clogging in the system piping at pipe reductions. It was assumed that the fibrous character of primary sludge caused the clogging. Grinding of the primary sludge in this current testing apparently prevented this problem.

## EMISSIONS AND POSSIBLE CONTROLS

Arrangements were made with the Air Quality Management District (South Coast Region) to monitor emissions while the plant was operating. Unfortunately, an emergency plant shutdown occurred on every scheduled date for measurement. Evaluation of the exhaust emissions from the pilot operation must be based solely on observations. The odor of ammonia from the process condensate collection system was always present, indicating uncontrolled venting. Throughout the testing there were uncontrollable losses of Isopar to the atmosphere from the mix tank, grinder, centrifuge, oil tank, and vacuum pump exhaust. These losses were not only observed, but also revealed in the mass balance calculations.

It is probably safe to state that even if pilot plant emissions had been monitored, the results would not have had direct application to a full-scale installation. Oil vapor emission discharges discussed above would be controlled in full-scale design. All sources of oil vapor emissions would be collected and the vent gases passed through a condensing unit for additional recovery of any residual oil. Remaining gases would be used as combustion air in a pyrolytic reactor or boiler, where they would be combusted under controlled conditions to reduce hydrocarbon emissions. Potential ammonia discharges for the high pH condensate could probably be controlled by direct recycling and dispersing into the treatment plant flow. This would reduce the pH and convert most of the ammonia to the ionized form which would not volatilize from solution.

## PATHOGEN SURVIVAL

Samples of the feed sludge, centrifuged solids, and hydroextracted solids were tested by the Los Angeles County Sanitation Districts' San Jose Creek Water Quality Laboratory for pathogen survival. The results of the analysis should be taken as tentative since only two sets of samples were analyzed. In general, pathogen concentrations were reduced by three log units from the sludge feed to the centrifuged solids. Further reduction might be expected in a multiple-effect system because of the longer contact time, and the reduced potential for short-circuiting. Pathogen concentrations in the hydroextracted solids indicate that this process may be effective in pathogen destruction. It seems logical to expect that the dry sludge resulting from the Carver-Greenfield process would present little or no public health concerns from biological agents in regard to the ultimate disposal method selected. The data is shown in Table 2.12.

## TABLE 2.12

## SURVIVAL OF INDICATOR AND PATHOGENIC MICRO-ORGANISMS DURING THE CARVER-GREENFIELD PROCESS

| Sample | Bacteria MPN/g Dry Weight | | | | Ascaris Ova | | Virus a/ Isolation |
|---|---|---|---|---|---|---|---|
| | Total Coliform | Fecal Coliform | Fecal Strep. | Salmonella | Total Ova/g | Percent Viable | |
| Feed | $2.8 \times 10^7$ | $2.8 \times 10^7$ | $2.7 \times 10^6$ | $5.0 \times 10^4$ | 28 | 33 | 7+/7 |
| Centrifuged Cake | $1.4 \times 10^4$ | $1.4 \times 10^4$ | $6.9 \times 10^2$ | <0.3 | b/ | -- | b/ |
| Dry Sludge | 0.7 | 0.7 | 2 | <0.2 | <0.2 | -- | 1+/8 |

a/ Number of positive isolations per number of tubes inoculated.

b/ Not tested - oil in sample not compatible with concentration.

## ENERGY BALANCES

Initial attempts to obtain a system energy balance were unproductive. Unaccounted heat losses from the process components were too numerous and indeterminable. Components where heat losses would be critical, such as the heat exchanger, were not insulated. The pilot plant was not engineered to demonstrate the heat efficiency of the process, even in the single-stage mode.

It was decided that, as an indication of the process heat efficiency, the pounds of steam applied (collected as a condensate) would be compared with the pounds of process condensate. Under the most favorable thermodynamic conditions for a single-effect evaporation process, one pound of steam should produce a little less than one pound of water (condensate), a ratio of close to 1:1 (steam:condensate). Field results are shown in Table 2.13. In all cases the ratios were 3:1 or greater. The increasing ratios with each phase of testing seems to reflect the greater frequency of plant shut-down due to equipment malfunction. In Phase 4 the data was not considered reliable because of the discontinuity of the test runs. Heat efficiency data from this pilot operation appears to have little application to a full-scale design.

## MATERIAL BALANCES

Results of material balance inventories are summarized in Table 2.14. Values in Table 2.14 represent calculated percent weight losses or gains of solids, water and oil through the pilot plant, for each phase of testing. An ideal mass balance would show zero loss or gain of material through the system. The actual results of the inventory would seem to indicate only marginal success at achieving a total material balance.

The material balances included all system inputs and outputs, as well as all material stored in the process units at the end of each phase. As mentioned previously in this report, the inventory of material fed into the system included: sludge feed, make-up Isopar, and initial dry charge material for process start-up. Material discharged from the system included: process condensate, centrifuged solids, centrifuged oil, various spills, and estimated evaporation losses from open tanks and other uncontrolled points.

Discrepancies in the material balances were due to three known causes: occasional improper sampling and inaccurate laboratory analysis; equipment malfunction which necessitated that the plant be operated on idle or even shutdown briefly; and weight measurement calibration errors. The first two were virtually uncontrollable. The third was evaluated, and although some significant discrepancies were found between some scales, major errors were self-cancelling. The magnitude of any calibration error which would be ultimately introduced into the material balance calculations was deemed minor.

## TABLE 2.13

## PROCESS HEAT EFFICIENCIES

| Phase | Steam Applied lbs. | Process Condensate lbs. | Heat Efficiency Steam:Condensate |
|---|---|---|---|
| 1 | 30,900 | 10,900 | 2.8 |
| 1A | 22,100 | 6,200 | 3.6 |
| 2 | 16,300 | 2,800 | 5.8 |
| 3 | 10,900 | 1,600 | 6.8 |
| 4 | Data not reliable | | |

## TABLE 2.14

## RESULTS OF MASS BALANCE INVENTORY
Percent Loss or Gain Through Pilot Plant

| Phase | Solids | Water | Oil |
|---|---|---|---|
| 1 | - 32% | - 5% | - 24% |
| 1A | + 27% | + 3% | + 11% |
| 2 | - 20% | - 15% | - 29% |
| 3 | - 36% | - 26% | - 72% |
| 4 | Data not available | | |

## OPERATING PROBLEMS

Operating problems and equipment malfunctions are defined in this report as process problems, and equipment problems, respectively. This was done in order to distinguish between problems related to the process itself and which, therefore, would be expected in a large scale system, from equipment problems on the pilot plant, which may not be related to full scale operation.

Although agitators were employed for the mix tank and the feed tanks, there was some difficulty in keeping the solids in suspension with the thicker dewatered digested primary sludge. The problem was solved by recycling a portion of the final solids-oil slurry to the mix tank. This apparently increases the solids to water ratio just enough to retard settling of the solids.

Foamover was experienced from the vapor chamber into the surface condenser, causing solids to contaminate the process condensate. This problem began in Phase 1A, and occurred intermittently throughout the remainder of testing. The cause was diagnosed by Dehydrotech as a cavitating recirculation pump. Modifications were made to the pump, but this provided only a partial solution. On the recommendation of Dehydrotech, small quantities of an emulsifier were added to the mix tank to alleviate the problem when needed. This kept the problem under control. An ultimate solution would probably involve redesign of the pilot plant hydraulics. The cause of foamover in this study was considered to be related to the pilot plant design, and only indirectly to the process itself or to the characteristics of the sludges tested.

Plugging problems were encountered in the addback and centrifuged feed tank lines, both of which carry the system discharge, or final solid-oil slurry. It occurred occassionally when there was a brief interruption of flow due to an equipment malfunction. The only solution was to disassemble the plugged line and force the contents out. Plugging of sludge slurry lines in full-scale installations has not been reported as a problem, apparently the result of proper oil:solids ratios, reliable continuous flow operation, and properly sized pipelines.

Formation of a gummy phase was considered to be a potential problem with the liquid digested primary and the thickened waste activated sludges. The sludge/oil mixture in a Carver-Greenfield evaporation system becomes sticky, or gummy, at certain solids concentrations; that is, when a certain amount of water has evaporated off. This concentration occurred in the pilot system only when dilute sludge feeds were batch dried prior to the initial runs so the system could begin with an equilibrium concentration. Subsequent runs utilized product solids from the previous run. After initial batch drying of liquid digested primary sludge, the entire system was purged. Because formation of a gummy phase was not detected at that time, the system was not purged after batch drying of the thickened waste activated sludge. The system was not inspected or cleaned again until after all testing had been completed. As mentioned elsewhere in this report, no fouling or clogging of the heat exchanger tubes was noted when the plant was disassembled. The gummy phase ordinarily affects only the heat exchanger, when it occurs.

## EQUIPMENT MALFUNCTION

Equipment malfunctions leading to system failures were experienced throughout the pilot plant operation. Shakedown of the pilot system took over a month. Between Phases 1A and 2, extensive modification to the hydraulic drive system necessitated a shutdown for two weeks. Pump seals and motors frequently failed. Sludge feed agitators and the sludge grinder malfunctioned occassionally. The problems became so numerous toward the end of testing that pilot plant operations were suspended early.

Analysis for the problems caused by equipment malfunction indicated strongly that the reliability of the process was not involved. The 1975 CLA pilot study (5) was conducted on essentially the same equipment, and was virtually free of mechanical difficulties. Apparently, the pilot unit equipment and design are geared to short term test runs. Operation under long term, continuous conditions apparently taxed the equipment beyond its intended design.

## PILOT PLANT LIMITATIONS

Some important aspects of the Carver-Greenfield process can only be demonstrated by a full-scale system. The pilot plant testing succeeded in proving a number of vital features of the Carver-Greenfield process, but severe unknowns still exist.

A detailed energy balance is among the most crucial. The heat efficiency of the system should be more accurately demonstrated. Actual energy saving benefits of multiple-effect evaporation is pivotal in deciding the cost-effectiveness of the system. The energy balance must include all ancillary systems such as hydroextraction and heavy oil separation.

Potential losses of Isopar from the system and the corresponding requirement for make-up oil must be determined. A totally enclosed system with oil recovery and recirculation units can be the only basis for judgement. A significantly higher than expected amount of Isopar loss in a full-scale system could alter the economic feasibility of the process dramatically.

Potential problems related to the build-up of fine solids in the recirculated fluidizing oil must be determined. Degradation of the fluidizing oil may necessitate an additional polishing process, or may otherwise render the process less desirable.

Potential sources of air pollutants must be evaluated. Stringent air quality regulations in the Los Angeles metropolitan basin are foreseen as a crucial hurdle to acceptance of the Carver-Greenfield and pyrolysis system by regulatory agencies.

Odors generated from the system are not well defined. A full-scale design would provide the basis for assessing the potential sources of odor and the methods for containment. Sources of odor experienced at this pilot facility do not relate too closely to any major installation.

Information obtained from the pilot operation contributed only indirectly to developing capital and operating cost data for a full-scale system. Pilot operation revealed the need for certain important components including control and instrumentation, hydroextraction, hydrocarbon ventilation system, and so on. The pilot operation also revealed certain operating characteristics such as fluidizing oil in the hydroextracted solids, etc. However, accurate cost data were not extrapolated from the pilot unit.

## CONCLUSIONS

The conclusions derived from the pilot plant demonstration are given below. The stated test objectives are shown in conjunction with the corresponding conclusions.

Objective 1:

To identify operational problems encountered in the continuous flow system.

Conclusion:

The operating problems encountered during the pilot-plant investigation were: solids settling out in the mixing (fluidizing) tank and feed tanks; foamover from the vapor chamber to the surface condenser; plugging of sludge-oil slurry lines; and dust contamination of the hydroextracted oil. Such problems were attributed to the shortcomings of the pilot-plant design and not the Carver-Greenfield process itself.

Objectives 2 and 3:

To develop design and operating parameters for a full-scale system;
To obtain capital and operating cost data for a full-scale system;

Conclusion:

Some basic design and operational data were developed from the pilot-scale investigation. However, a large scale plant (30,000-60,000 lbs/hour of evaporative capacity) is needed to be built and operated for generating reliable design and operational data. This would also produce meaningful capital and operating cost data for a full-scale plant.

Objective 4:

To investigate the efficiency of the hydroextraction process for removal of residual oil in the dewatered solids;

Conclusion:

The hydroextraction process was effective in removing residual oil in the centrifuged solids. Hydroextracted solids contained 1.1 percent oil or less (based on dry weight of solids).

Objective 5:

To identify corrosion problems;

Conclusion:

No scaling of the heat exchanger tube inner walls was observed. It appeared that Isopar L was an effective fluidizing medium. During investigation no corrosion problem was noticed.

Objective 6:

To determine the extent of heat exchanger tube fouling during continuous operation and the ability of sludge grinding to mitigate the problem;

Conclusion:

Clogging of the heat exchanger tube nest inlet by fibrous sludge material was prevented by disintegration (grinding) of the sludge feed. Primary sludges definitely require disintegration; waste activated sludges may not.

Objective 7:

To determine the characteristics of final products resulting from the system including liquid sidestreams (i.e., condensate) and exhaust gases;

Conclusion:

a. Extracted sludge water (process condensate) contained TDS between 16 and 22 mg/l, COD between 212 and 4167 mg/l, nitrogen as ammonia between 464 and 1749 mg/l, and heavy metal concentrations generally below 0.1 mg/l.
b. Dried sludge solids (centrifuged solids) contained residual oil ranging between 37.2 and 47.4 percent, total solids between 51.3 and 60.9 percent, and moisture between 1.0 and 7.7 percent. On an oil-free basis, the solids concentration ranged between 87.2 and 98.1 percent, with an average value of 95.7 percent.
c. Recirculated oil became contaminated with sludge oils and fine sludge solids. Stripping of the recirculated oil is effective in removing both heavy sludge oils and sludge solids, and would help control the inventory of fine solids in the system.
d. Based on limited data, pathogen destruction through the Carver-Greenfield process appeared to be effective.
e. Stripping of the fluidizing oil may effectively separate fluidizing oil from the heavy oils and should aid in preventing excessive build-up of heavy oils in the system.
f. System losses of Isopar were due primarily to uncontrolled venting. If vent losses are controlled, potential loss of Isopar should be limited to that contained in the final hydroextracted solids.
g. Major emissions of concern with the Carver-Greenfield process itself are likely to be potential hydrocarbon emissions resulting from the volatility of the Isopar. Mix tanks, vacuum pumps, centrifuge, and solids conveying systems were identified as the major sources of such emissions. In a full scale system, provision must be made to collect these vapors, condense them, and direct remaining gases to a boiler or pyrolysis reactor for combustion.

Objective 8:

To develop mass balances for the system of Carver-Greenfield hydroextraction and pyrolysis including fluxes of solids, water and oil;

Conclusion:

Valid energy balances could not be obtained because the pilot plant was not engineered to minimize heat losses.

Objective 9:

To develop energy balances for the one-effect system including hydroextraction and pyrolysis and extrapolation of these balances to a multiple-effect system.

Conclusion:

No meaningful mass balances could be obtained from the pilot study.

# CHAPTER 3

## CHAPTER 3

## REVIEW OF EXISTING CARVER-GREENFIELD INSTALLATIONS

Although there are over sixty operating Carver-Greenfield (C-G) installations, they are mostly used for processing industrial materials. The installations processing municipal sewage sludge or materials similar in nature are listed in Table 3.1. To develop first-hand knowledge of the operating experiences from full-scale plants it was decided by the LA/OMA Project to visit certain of these installations. To date two of the plants, one operating (Coors) and the other shut down (Omaha), have been visited by the LA/OMA Project staff. The reports of the field trips are given below.

### SITE VISIT TO
### C-G INSTALLATION AT ADOLPH COORS BREWERY
### GOLDEN, COLORADO

Dr. Roger Haug, Senior Engineer, LA/OMA Project, Stan Naydo, Bureau of Engineering, City of Los Angeles and Mike Cooney, State Water Resources Control Board, State of California visited the C-G installation at the Adolph Coors Brewery, Golden, Colorado, on September 13, 1977.

The visiting party was escorted through the facility and held most of their discussion with Mr. Jack Bays, Manager of Process Development, Research and Development Department for the Adolph Coors Company. They also talked with Mr. Bob Mason, Manager of the Water and Wastewater facilities for the brewery. The following is an excerpt from Dr. Haug's trip report (11).

#### SYSTEM DESCRIPTION

The Carver-Greenfield process at the Coors Brewery is a 4 effect system sized for an evaporative burden of 60,000 lbs/hr which equates to 30 dtpd (dry tons per day) at 4 percent solids. Current waste activated sludge production at the wastewater treatment plant averages about 14 dry tons per day. At the present time, the Carver-Greenfield process is run at a rate of 30 dry tons per day on an average of 3 days per week. The evaporation unit was oversized because brewery production is expected to increase from 15 million barrels of beer per year to about 27 million barrels per year within the next 8-10 years. This will significantly increase the wastewater sludge production and it was felt to be more cost-effective to provide the extra capacity initially rather than expand the facilities at a later date.

The largest single component in the combined flow to the wastewater treatment plant is from the malting process, which contributes about 40 percent of the total flow. The aging process and house cleaning provide most of the remaining wastewater flow. Total COD in the plant influent is approximately 1200 mg/l. About 15 percent of this COD is suspended material averaging about 200 mg/l. Nitrogen in the form of ammonia is added to the wastewater because of nutrient deficiencies. Primary treatment facilities are available at the treatment plant. On the day of our site visit, however, these facilities were not being used and the raw wastewater was being added directly to the activated sludge system. Mixed liquor suspended solids are maintained between 2000-2500 mg/l. Return waste activated sludge is reaerated for an estimated period of 6-9 hours before being blended back into the raw wastewater flow. After blending, about 16 hours of aeration are provided. Returned waste activated sludge averages between 0.5 to 1 percent solids. Waste activated sludge is dissolved air floated to about 4 percent solids and stored prior to processing in the Carver-Greenfield system. Visual observation of the mixed liquor indicated that in all respects it was similar to waste activated sludge produced in municipal secondary treatment facilities.

At the present time (September, 1977) less than one percent of the flow to the wastewater plant is from domestic origin. This fraction is being removed from the system so that the Carver-Greenfield product can be sold as an animal feed supplement. The Food and Drug Administration (FDA) has ruled that the dried product cannot be sold as an animal feed supplement until the sanitary wastes are removed. Therefore, the Carver-Greenfield product is currently being landfilled. Removal of the sanitary wastes is anticipated by summer of 1978. The material will then be marketed as a vitamin $B_{12}$ supplement for animal feed. According to Coors staff, FDA specifies that the $B_{12}$ content must be greater than about 1.5 milligrams per pound in order to be classified as a supplement. Coors sludge currently runs about 3 milligrams per pound. To my knowledge this will be the first fullscale commercial application where waste treatment sludge is converted to an animal feed.

Pathogen destruction in the Carver-Greenfield process has been monitored. Pathogen counts in the final product were noted as being "good". When further questioned on this matter it was indicated that it is not uncommon to find low counts of fecal strep and other indicator organisms in the final product. It is not known at this time whether these low count in the final product are due to recontamination from the atmosphere or residual carryover through the process. Temperature in the last stage is nearly 250°F, and it seems unlikely that pathogenic organisms would survive under these conditions. However, each effect is probably completely mixed and therefore considerable material could pass through a single effect in less than the theroretical detention time. With four effects in series, however, short circuting through the entire system would be minimized. Further clarification of the pathogen destruction in the Carver-Greenfield system is needed.

The Coors facility started operation in the fall of 1976. It operated approximately 60 percent of the time until February, 1977. By that time, start-up problems had been overcome. Since

February, the unit has been run on a regular basis, 3 days per week. A total of about 4,000 dry tons of solids has been processed in the system since start-up.

During the initial start-up of the Carver-Greenfield system plugging problems were encountered. This was due to formation of a gummy phase which was found to occur at a certain ratio of solids to water. Possible formation of such a gummy phase was previously described to us by Dehydrotech in conjunction with the LA/OMA pilot study. Mr. Bays thought that the gummy phase was due to a phase inversion. In other words, conversion from a system where water is the continuous phase to a system where oil is the continuous phase. Assuming oil and solids are mixed in a ratio of about 6:1 by weight, the initial mixture would contain about 4 part water by weight per part of oil, assuming 4 percent sludge solids. Water would certainly be the continuous phase in the initial mixture. As water is evaporated, however, oil must eventually become the continuous phase at some point in the system. While this explanation is plausible it should be viewed as preliminary.

Fortunately, the gummy phase occurs only at a certain ratio of solids to water. The Dehydrotech people were well aware of the problem according to Mr. Bays and easily overcame it when it was noted. Since about one-fourth of the total water input to the system will be evaporated in each of the four effects, each effect will operate at a certain solids to water ratio. Therefore, it is entirely possible that the gummy phase will be avoided in each of the effects. If this should not be the case, addback of dry slurry from the 4th stage to the fluidizing tank could be used to overcome the problem. Similar information was previously supplied by Dehydrotech. Further clarification is needed on the nature of the gummy phase and the effectiveness of addback in controlling it.

Subsequent conversations with Mr. Greenfield on September 23, 1977, further clarified the gummy phase condition and the effect of addback in controlling it. Apparently, the gummy phase can occur at both a low and high ratio of solids to water. The actual ratios depend on the sludge type, carrying oil, and other factors. Typical ratios for municipal sludges using petroleum base carrying oils might be 30 percent solids on the low end and about 70 percent solids on the high end. The lower range can be avoided by addback of dry solids and carrying oil from the 4th stage to the fluidizing tank. If the gummy phase occurred at 30 percent solids, recycle or addback would be adjusted to provide at least 30 percent solids in the fluidizing tank. It was Mr. Greenfield's opinion that in a commercial scale plant, addback could probably be adjusted to achieve only a 25 percent mixture because of the greater mixing and shear stress provided in a full scale system. Heat energy contained in the recycled stream would be recovered by heat exchange with the vapor chamber fluid as it is transferred from stage to stage.

An alternative to the use of addback is the addition of an emulsifying agent to the carrying oil. In a commercial whey plant about 0.2 percent emulsifier is added to the carrying oil for control of the gummy phase. However, Mr. Greenfield indicated that Dehydrotech has more experience with the addback feature and places more confidence in it.

A gummy or sticky phase can also occur in the 4th stage of the system if the ratio of solids to water is too low. If solids in the 4th effect are too wet, the oil can fall away from the mixture allowing the solids to stick to the exchanger tubes. This can be avoided by controlling the feed to the last effect to assure that the input solids are not too wet.

From the above discussion, it is obvious that the Carver-Greenfield process must operate within certain limits or boundaries and that the system must be tuned to avoid these limiting conditions. Once the boundaries are established, reliable operation can apparently be achieved as evidenced by the experience at Coors.

Mr. Bays indicated that they originally had planned to run the unit continuously, seven days a week, at a reduced loading rate because of concern over start-up and coordination of the four effects proved to be easier than anticipated. The system can now be "lined out" in about 6 hours. Therefore, a procedure was adopted whereby sludge is stored and run through the Carver-Greenfield at full capacity for 2½-3 days a week. This means that the system is started up and shut down once each week.

The unit was stated to operate at about 90 percent of theoretical efficiency. In other words, only about 10 percent of the input energy is lost through radiation and other losses. Foam insulation was applied to both heat exchanger and vacuum chamber surfaces. Qualitatively, it was observed that radiation losses were minimal since one could stand next to the vapor chambers and even touch them without experiencing significant heat.

During the course of the site visit, Mr. Bays quoted a number of different thermal steam efficiencies. In our morning meeting, it was indicated that the unit processed about 3.6 pounds of water per pound of steam applied. In the afternoon, a value of 3.9 pounds of water per pound of 400 psi steam was indicated. The difference between these numbers may be in the pressure of the applied steam. During the afternoon, I spent about one-half hour in the control room and noted a steam supply of 20,000 pounds per hour corresponding to about 115 gallons per minute of condensate flow. This would correspond to about 2.9 pounds of water per pound of steam. However, there is no way of knowing whether the meter calibrations were accurate. In any event, it is likely that the thermal efficiency of the system is no less than 3 pounds of water per pound of steam and quite likely higher.

The recirculating pumps, vacuum chambers and heat exchangers occupied a space about 30 feet wide by 80 feet long. Height from the recirculating pumps to the top of the heat exchangers was estimated to be about 100 to 120 feet. If a space of only 30 by 80 foot is required, the recirculating pumps and barometric leg, which account for 30 to 40 feet of the total height, could be placed underground.

Heat exchangers, heat exchanger tubes, vacuum chambers, and associated piping are constructed of 304 stainless steel. Mr. Bays indicated that in his opinion it was not necessary for the vacuum chamber or piping to be of stainless steel. It was used in the Coors facility because the brewery is set up to use stainless steel and even has its own stainless steel fabricating plant. He felt that construction of the vapor chambers and piping out of mild steel would be adequate. This corresponds to information previously submitted to LA/OMA by Dehydrotech, which recommended the use of 316 stainless steel in the heat exchangers, using 1½-inch O.D. tubes, 18 gauge in all stages except the last which would be 16 gauge. Carbon steel was recommended for use in the vapor chamber and piping.

Flash pots, centrifuge, vacuum pump exhaust, and other points in the system are vented to collect exhaust emissions and control possible odors. The vent gases are taken to a small incinerator which is fired with natural gas and combusted at about 1400°F. Considerable displeasure was noted by Mr. Bays over the incineration system because of its large energy consumption. He indicated that it was his desire to install a scrubbing system for odor control in the vent gas. I indicated that we have proposed to pass the vent gas through a condensing unit to recover remaining oil and then use the gas into a steam boiler as combustion air. He indicated that this was a good idea and could see no reason why it should not work.

Condensate from each effect is combined and passed through an API separator for oil/water separation. The API separator is completely enclosed and vented. Product water from the bottom of the separator is passed through a sock filter which removes residual oil and then through a coalescer. The latter is an enclosed unit packed with a porous plastic bed. Emulsified oils cling to the surfaces of the bed, coalesce, and then eventually break from the bed and rise to the surface where they are removed and returned to the system. Final condensate water from the coalescer has averaged between 2000 and 4800 mg/l COD, 100 mg/l ammonia-nitrogen and few mg/l of carrying oil. The water is recycled to the activated sludge treatment plant.

Concrete flooring with an epoxy covering was provided on all floors of the building for ease in cleaning up unavoidable oil spills. Wash water at 600 psi and 180°F is supplied and is indicated to be very effective in cleaning off oil from the concrete flooring. It was advised that such a washwater system should be provided in any facility which uses oil as a carrying media. Washwater and seal water from the bearings of the pumps has averaged about 20 gallons per minute. Capacity should be provided to feed this water back to the system, probably to the oil-water separator used to treat the condensate water. The high degree of cleanliness observed in the Carver-Greenfield building is undoubtedly due in part to the philosophy of cleanliness necessary in the brewery operation.

Tallow, an animal fat, is currently being used as a carrying oil. The system was first started using a petroleum based oil and then converted over to the animal based oil. The use of an animal or vegetable based oil is necessary in the Coors system because of the desire to market the final product as an animal feed substitute. The oil to solids ratio in the blended material fed to the first effect is approximately 6 parts of oil to one part dry solids according to Mr. Mason. Mr. Bays stated that the ratio was in the range of 10:1 to 15:1. The discrepancy may be due to the fact that correction factors, developed during instrument calibration, are applied to the gauge readings in the control room. Further clarification of the exact oil:solids ratio is needed.

Carrying oil and sludge from the last effect is fed to a Bird solid bowl centrifuge. The Bird machine was of modest size appearing to have a scroll about 24 inches in diameter by 6 feet long. Cake solids from the machine have averaged about 35 percent solids. Assuming 120 gpm of sludge flow at 4 percent solids and an oil to solids ratio of 6:1, flowrate to the centrifuge was calculated to be about 40 gpm and would consist of about 14 percent solids and 86 percent oil by weight. This agrees closely with a stated centrifuge feed consisting of 15 percent solids, 3 percent water, and 82 percent oil.

Solids content in centrate oil from the Bird centrifuge was stated to be in the range of 2 to 3 percent solids. Assuming cake solids of 35 percent and oil:solids ratios of 6:1 and 10:1, calculated capture efficiencies would be 91 and 83 percent, respectively. While these are somewhat lower removal efficiencies than would be desireable for conventional water/solids separation, they do not appear to pose any significant problems in the Carver-Greenfield system. Mr. Mason indicated that the high solids content of centrate oil was not of serious concern and no build-up of fine solids within the Carver-Greenfield system was observed. This is a point which should be further explored. Hopefully, the LA/OMA studies conducted at Hyperion will shed more light in this area.

It is interesting to note that the high solids content in the centrifuged oil is not directly comparable with desired centrate qualities in water:solids separation. Basically, this is because feed to the Bird machine is about 15 percent solids which is quite high compared to the thin slurries normally fed to dewatering equipment in municipal applications.

The following specifications for pumps and other ancillary equipment were obtained.

Vacuum pumps - Two units each of 25 horsepower. Both units are operated during start-up to pull down pressure in the system. During study state operation, one vacuum pump is in operation with one stand-by.

Circulating Pumps - One for each effect, each pump 100 horsepower Waukesha lobe type pump, constant speed.

Transfer Pumps - One per effect, 5 horsepower each, constant speed.

Centrifuge - The drive unit was not checked but it was estimated to be about 100 horsepower.

These values for installed horsepower are in close agreement with values projected for LA/OMA by Dehydrotech. Note that all pumps are constant speed with flow control provided by control valves.

Regarding stand-by equipment, one spare recirculating pump and one spare drive unit is maintained in the building in addition to the one stand-by vacuum pump which is not used in steady state operation.

A macerator or grinder is not used on the system to grind the sludge prior to entry into the Carver-Greenfield. Although some primary sludge can be produced in the treatment plant, the predominant waste activated sludge nature of the material probably lessens the need for any grinding of the sludge. Maceration of the sludge was determined to be necessary based upon preliminary LA/OMA pilot plant results, at least with primary digested sludge.

The required heat transfer surface in each of the heat exchangers varies somewhat. U rates in each effect are slightly different because of the varying nature of the material and the different temperature drops maintained across the heat exchangers. Tubes in the heat exchangers are all 1.25 inches inside diameter. It was pointed out that oversizing the heat exchangers does not affect the operation of the Carver-Greenfield process. However, undersizing of the heat exchangers would cause problems in operation of any multiple effect system. Therefore, it is most important to assure that each effect has sufficient heat transfer surface.

Plant effluent is used for cooling water in the final condenser. No problems of corrosion or fouling by biological growth have been noted. I suspect that this effluent is chlorinated prior to being used in the exchanger, however.

## CAPITAL COSTS

The original cost estimate for the Carver-Greenfield installation at Coors, including building costs, installation and all equipment cost, was 2.5 million dollars in mid-1973. Coors management delayed a final decision regarding installation of the Carver-Greenfield system. When the decision was finally made, actual cost for the complete facility was 5 million dollars corresponding to about March, 1976. The latter cost does not include costs for a boiler facility since in-plant steam is used in the process. The cost does include the Carver-Greenfield evaporation system, installation, building costs, pumps, dewatering equipment, condensate treatment system, all associated piping and other ancillary equipment.

It should be noted that the building which houses the Carver-Greenfield is quite spacious and could probably contain one or two additional modules of the same size as the existing system. Therefore, capital cost for a multi-modular system may be significantly less than that predicted from extrapolation from the Coors data.

## OPERATION AND MAINTENANCE COSTS

### Labor

Labor requirements for the Carver-Greenfield include about 1.5 operators per shift plus an average of about 20 man-hours per week maintenance. Regarding supervision, the operators and maintenance personnel are under normal plant supervision. No additional supervisory staff were employed as a result of the Carver-Greenfield system. There are no shift foremen and the Carver-Greenfield is under the direct control of the operator. An engineer is on site only during the day shift. Operators for the CG unit were employed from within the treatment plant section of the Coors brewery with some individuals being hired from outside. It was indicated that these individuals had no special or unusual talents prior to their assignment to the Carver-Greenfield. Coors engineering staff conducted a training program for the operators prior to start-up of the Carver-Greenfield.

### Chemicals

Oil content of the final product at Coors ranges from 10 to 20 percent which equates to about 6000 to 12,000 lbs/day of tallow assuming 30 dtpd of sludge processed. The price of tallow oil fluctuates widely but averages about $.16/lb. This calculates to about 32 to 64 dollars per ton of dry sludge solids, depending on the final oil content. Obviously, with municipal sludge, which likely cannot be sold as animal feed, such costs would be prohibitive. This seems to substantiate the decision to concentrate on use of a steam extractable carrying oil and the use of hydroextraction in place of screw pressing.

In addition to the tallow oil, a chemical called Twen 80 is used as an emulsifier. This chemical is not used at all times but when used was stated to add about $2.00 per dry ton to the operating costs. Antifoam can also be added to the system but is not used on regular basis. It is interesting to note that a moderate foaming problem was observed in the activated sludge system. However, no carryover of this foaming problem into the Carver-Greenfield system was noted.

## Maintenance

No major equipment problems have been noted in the evaporator system itself and, at this time, there is no fixed schedule of major maintenance or cleaning on the heat exchangers. In the opinion of Mr. Bays, routing of the heat exchangers every six months to one year would be adequate. Complete retubing of each heat exchanger once every ten years should also be expected.

Some erosion problems have been noted in the Bird centrifuge, and the screw press has been out of service on several occasions because of mechanical difficulties. Use of ceramics on the centrifuge scroll is being investigated. It seems unlikely that wear on the centrifuge would be more than normally encountered in centrifuging a solids:water system.

## OTHER COMMENTS

I questioned Mr. Bays regarding analysis of alternative systems prior to the decision to implement the Carver-Greenfield process. He indicated that Coors looked at several alternatives including land application of the sludge. Mr. Bays conducted extensive testing of the Carver-Greenfield process prior to the decision to implement. Tests were run using the CG installation at the Alpo plant in Allentown, PA, and also at the Carver-Greenfield facility in New Jersey. The CG installation in Hershey, PA, was also visited. Considering the specifics of their situation, the Carver-Greenfield system was determined to be the most attractive alternative.

Regarding the Coors decision to produce an animal feed from the waste sludge, several other animal feed materials are produced at the brewery and marketed by Coors. Spent grain from mashing is dewatered, dried, and sold as a cattle feed supplement. Excess yeast from fermentation is separated and used as a poultry feed. In addition, a double effect evaporator is used for concentrating waste beer and spent grain liquor into cattle feed syrup. In light of these existing marketing efforts, the decision to produce a $B_{12}$ supplement from the waste sludge is understandable.

## SITE VISIT TO
## C-G INSTALLATION IN OMAHA, NEBRASKA
## OCTOBER 6, 1977

In 1969 a Carver-Greenfield (C-G) plant was built in Omaha for processing packinghouse wastes. The plant started in early 1970 and had to be shut down almost immediately for major operating problems. Since then the Omaha experience has been cited by many as a classic example of the C-G processes inability to be adapted in the sludge handling operations. Therefore, when the LA/OMA Project undertook an indepth study of the C-G process, a decision was made to investigate the Omaha incident. Sagar Raksit, Project Engineer, LA/OMA Project made a site visit on October 6, 1977 (12).

Mr. Charles Geisler, Sanitation Engineer for the City of Omaha took him to the abandoned C-G plant and explained the problems experienced during the short operation life of the installation. Mr. Geisler was very helpful in explaining and discussing various aspects of the problems they faced, and offered his views on the reasons for failure of the system. The following description summarizes the information obtained from Mr. Geisler as well as from observations made during the plant tour.

### SYSTEM DESCRIPTION

On December 1, 1966 Carver-Greenfield Corporation (CGC) entered into a contract with the Omaha Pollution Control Corporation (OPCC) to engineer, construct and operate a plant for treating wastewater from meat packing companies located in the City of Omaha. These meat packing houses processed on an average of 2 million beef cattle per year as well as an appreciable load of hog and sheep. The wastewater thus generated contained 180 tons/day of paunch and 55 tons/day of grease (animal fat). The basic aim of the treatment scheme was to recover the paunch and grease for sale. The process train included gravity settling of paunch followed by Carver-Greenfield (C-G) process for recovering grease. CGC also agreed under contract to meet certain specific composition for the reclaimed grease.

Prior to the execution of said agreement, the City of Omaha hired Stearns-Roger Corporation of Denver, Colorado, as consultant to investigate and study the feasibility of the Carver-Greenfield proposal. In their report to the City, they concluded that the proposal was economically, commercially and technically feasible. The preliminary testing data was gathered through tests conducted at a sewage plant located in St. Joseph, Missouri, since it was assumed that the materials processed in that location would be similar to those expected in Omaha.

The construction of the facility was substantially completed around November, 1969. Plant shake down process was hindered due to an unusually severe winter. Start-up efforts began in March 1970 and almost immediately the plant experienced two operational difficulties. The first concerned an oil-soluble soap buildup in the grease being produced in the plant which affected product quality and processing operations. The second related to heavy fines contained in the grease which also affected product quality and processing operations to a lesser degree. As a result, the final grease product could not meet the required quality specification which heavily affected its sale price.

In order to eliminate these unanticipated problems, CGC recommended that a permanent, continuous acidulation system be installed in the plant for the purpose of minimizing the soap buildup and it also suggested the replacement of a DeLaval centrifuge which had been incorrectly furnished by DeLaval Company and which would have reduced or eliminated the problem caused by the fines. The acidification of the feed material was demonstrated to be successful in alleviating the buildup problem due to soap. However, to incorporate a permanent acidulation step in the process train required considerable cost. The construction was estimated to be approximately $80,000.

The City of Omaha, which took over the responsibility of the plant operation from OPCC, felt that any cost incurred in such modification should be paid by CGC. However, CGC felt they were not obligated to bear such additional cost.

## RESULT

In August, 1970, CGC was declared to be in default under its contract. CGC contended that if the recommendations it had made had been approved and authorized, the plant would have produced sufficient revenues to satisfy the debt service out of the fund for the project and created a substantial profit for the City of Omaha. Following the alleged default, in November, 1970, the City of Omaha and the Omaha Pollution Control Corporation instituted a suit against CGC for damages. The trial court ruled that the failure to place the plant on stream constituted an act of default on the part of CGC.

## CONCLUSIONS

Following the site visit and discussion with both Mr. Charles Geisler (City of Omaha) and Mr. Charles Greenfield (CGC) the LA/OMA Project engineer has drawn the following conclusions.

1. The shutdown of the Omaha C-G installation could have been avoided by proper process modifications. It should not be termed as a process failure due to inherent process inadequacies but rather poor design or poor planning or both. A thorough waste characteristics study would have indicated the presence of oil-soluble soap in the wastestream which in turn would indicate the need for acidulation step. On the other hand, if such characteristics study was not possible before designing the plant, enough contingencies or redundancies should have been provided for future modification.

2. The said C-G installation was designed for processing solids generated from treating meat packing waste which has significantly different characteristics from sewage sludge. Therefore, the experience obtained from one is not directly applicable to the other; it should only be used as a guide after proper extropolation.

3. The problems encountered at the Omaha facility appear to have been solvable from a technical standpoint. The impact of surface active agents should be recognized. However, their concentration in municipal sludge is expected to be low. Therefore, the City of Omaha's experience with their C-G installation should not be viewed as a negative factor against applications of C-G process for drying sewage sludge.

# CHAPTER 4

## CHAPTER 4

## PYROLYSIS OF THE CARVER-GREENFIELD PRODUCT

Drying of sludge by the Carver-Greenfield (C-G) system followed by incineration or pyrolysis appears to be an attractive combination of unit process. Dry sludge makes the energy recovery more efficient whereas the incineration or pyrolysis could supply the energy required to operate the C-G process. Therefore, most proposals on treating municipal sludge with the C-G process include an incinerator or pyrolysis reactor to recover the heat value of the dried product.

One of the objectives of the pilot plant study (reported in Chapter 2) was to determine the suitability of the dry sludge product as a feed material for energy recovery. The Dehydrotech Corporation arranged to run a small-scale pyrolysis test at their own expense using the dry sludge produced from the pilot-plant work. The tests were conducted by the Surface Division of Midland-Ross Corporation at Toledo, Ohio. A report describing the test results was submitted by Midland-Ross to Foster Wheeler Energy Corporation (Consultant to Dehydrotech)(13). A summary of the relevant information from this report is presented in the following pages.

### TEST DESCRIPTION

About 2000 pounds of dried sludge produced by the Carver-Greenfield pilot plant was shipped from Los Angeles to the Dehydrotech test facilities located at East Hanover, New Jersey. The sludge was de-oiled by hydroextraction and extruded into pellets, ½-inch in diameter and ½-inch long. The pellets were then sent to Midland-Ross' test facilities at Toledo, Ohio.

At Toledo, preliminary gravimetric, rate of pyrolysis and bomb calorimeter tests were conducted to establish a basis of information for the pilot plant test. Results from these tests are shown in Table 4-1. The pilot plant tests consisted of two runs in a vertical reactor (shaft furnace pyrolyzer) and two more runs in a rotary hearth furnace. One of the test runs involving rotary hearth furnace was witnessed by Sagar Raksit, LA/OMA Project Engineer. The test descriptions and results involving the respective furnaces are given separately.

### TESTS INVOLVING VERTICAL REACTOR

The furnace used was a vertical shaft pyrolyzer with an inside diameter of 2 feet and height of 15½ feet. The furnace was equipped with an excess air burner and a Rich Fume Incinerator for combusting the fuel gas produced in the shaft furnace. The furnace was equipped with temperature controllers, temperature indicators, recorders, and proper flame and safety monitoring equipment.

A continuous feed mechanism was provided consisting of two main feed rams actuated by air cylinders, with a third air cylinder operating the furnace door. A 31 foot flight elevator was used to transport feed material from the floor level to the top of the shaft furnace. Loading of the pelletized material into the elevator was done manually.

**TABLE 4.1**

**FEED SLUDGE ANALYSES**

| | |
|---|---|
| Volatiles | 46.08% (by weight) |
| Fixed Carbon | 7.35% (by weight) |
| Ash | 46.57% (by weight) |
| Heating Value | 5152 BTU/LB |

Two experimental tests were conducted on the shaft furnace to determine its suitability in gasifying pelletized sewage sludge. In the first run, a reactor temperature of 1850°F and a bottom feed gas oxygen percentage of 17.5 were maintained. The sludge feed rate was 150 pounds per hour. The perforated plate around the air blast inlet at the bottom of the shaft furnace had ¼" diameter holes. The heat sources in the shaft furnace were the hot gases coming in from the excess air burner and the char combustion taking place inside the furnace.

It was observed that feed material fused and the pellets bonded together to form large pieces of hard rock-like structure. Very little ash came down through the perforated plate into the ash hopper. A possible reason for this, as suggested by Midland-Ross personnel, could be the presence of localized high-temperature spots inducing the slagging of some compounds which fused at this temperture. Another hypothesis was that the height of the bed in the shaft furnace may have contributed in causing pressure-bonding of the pellets at high temeperature.

In the second experimental run in the shaft furnace, the gasification temperature was reduced to 1500°F, inlet gas oxygen percentage was 11% and the feed rate, 100 pounds per hour. About 70% of the material was gasified but some of the material fused again and created problems. The reason for this fusion was thought to be non-homogenity of the feed material composition. Secondly, pressure-bonding due to the height of the bed could have been a factor. Also, even though the oxygen percentage was reduced, it could still cause localized high-temperature spots. Furthermore, reducing the feed gas oxygen percentage to still lower values, slows down the gasification process considerably, lowers the temperature and reduces the combustion of the char due to lack of oxygen. It was also concluded that by reducing the amount of oxygen fed to the furnace the char combustion reactions were severely handicapped and heat energy available from this source was, therefore, not released.

Because of the problems encountered with partial slagging and fusing of the feed pellets and because of the difficulty in gasifying or combusting the char at lower feed air rates and reactor temperatures, it was decided to discontinue use of the vertical shaft reactor in favor of a rotary hearth furnace.

## TESTS INVOLVING ROTARY HEARTH REACTOR

A rotary hearth pyrolyzer consists of a rotating hearth on which the feed material is distributed in from a half inch layer. Fixed rabble arms extend across the hearth. Rotation of the hearth causes the feed material to move across the hearth to the discharge point at the center of the hearth. The rotary hearth is analogous to a single hearth in a multiple hearth furnace. Therefore, the results can probably be reasonably extrapolated to the multiple hearth reactor as well as the rotary hearth.

The furnace used for the pilot tests had a hearth diameter of 2.5 feet with an effective hearth area of 2 $ft^2$. The heat for start-up was supplied by a throat-mix burner. The pyrolyzer was coupled with a rich fume incinerator for combusting the gases produced in the furnace.

Sludge was introduced in the furnace manually through a chute at a rate of about 15 lb/hr for a hearth area of 2 $ft^2$, giving a loading rate of 7.5 $lbs/hr/ft^2$ of hearth area. Sludge pellets, formed for use in the vertical shaft reactor, had to be crushed into smaller size particles before feeding. Char and ash was removed from the furnace by a curved scraper blade. Temperature in the furnace was maintained at 1500°F and the oxygen percentage at less than 1 percent. The rate of hearth rotation was kept constant at 1/6 r.p.m. Total residence time in the pyrolyzer was about 15 minutes.

The two test runs with the rotary hearth furnace were considered to be successful. Test results are shown in Table 4.2. Results from the rotary hearth tests indicated that Carver-Greenfield process produced a dry sludge capable of autogenous combustion without the need for supplemental fuel. Temperatures in the pyrolyzer were maintained without use of supplemental fuels. A heat balance conducted during the test runs indicated a heat release of about 11,150 BTU/lb of volatiles gasified which is reasonably consistent with calorimetric data in Table 4.1.

TABLE 4.2

ANALYSES OF CHAR FROM ROTARY HEARTH FURNACE

| | |
|---|---|
| Temperature | 1500°F |
| Volatiles | 2.7% (by weight of char) |
| Fixed Carbon | 14.7% (by weight of char) |
| Ash | 82.7% (by weight of char) |

Bench scale studies were conducted to determine the rate of pyrolysis for different sludge thickness on the hearth. Results are presented in Table 4.3. Based on this data the contractor felt that high hearth loadings greater than 7.5 lb/hr-ft$^2$ could be realized. However, no demonstration runs were conducted at higher hearth loadings.

## MAJOR OBSERVATIONS

1. Partial slagging and fusion of the feed pellets was observed in the vertical shaft reactor at a gasification temperature of 1850°F. Reducing the temperature to 1500°F reduced but did not eliminate the problem of fusion. Pressure-bonding due to the height of the bed and localized high temperature zones were proposed as possible explanations of the problem.

2. A rotary hearth furnace successfully gasified the volatile fraction of the feed material at a temperature of 1500°F. Slagging or fusion was not observed during the tests. Fixed carbon was not significantly gasified and was removed from the reactor in the ash. Material handling problems were not encountered during the rotary hearth tests.

3. A hearth loading of 7.5 lb/hr-ft$^2$ was found to be suitable in the demonstration runs. Bench scale testing indicated that higher loadings may be possible.

4. Pelletizing of the Carver-Greenfield dried sludge will likely not be required for gasification or combustion in hearth type reactors.

5. Weight of the char removed from the reactor was about half the weight of the input material. This would be reduced slightly if a higher percentage of the fixed carbon were gasified or combusted.

6. Fuel gas produced from gasification was combusted in an incinerator without auxiliary level. However, data on the gas composition and quality was not available.

# CHAPTER 5

# CHAPTER 5

## FULL SCALE SYSTEM CONCEPT

This chapter will discuss the potential role of the Carver-Greenfield process as part of a total sludge management system. It is difficult to evaluate a unit process without placing it in context with the other processes comprising a complete sludge management system. For example, any heat drying process no matter how efficient is energy intensive. However, if the dried product is thermally processed for energy recovery the combined system may be energy producing. Therefore, it is the purpose of this chapter to explore the Carver-Greenfield process as it relates to various management systems and to compare the latter systems with other "typical" management alternatives.

For a new plant whether anaerobic digestion would be a part of sludge management system or not will depend on the eventual disposition of the dried sludge product. If the dried sludge has to be landfilled, digestion seems to be highly desirable for recovering methane, stabilizing organics, and reducing the solids loading on the subsequent unit processes, including sludge hauling. On the other hand if dried sludge is to be used as a fertilizer, digestion may not be desirable in order to preserve the nitrogen content of raw sludge. If energy recovery through thermal processing is the goal of the system, the role of the digestion process has to be carefully evaluated. Energy available in sludges can be recovered either by digestion followed by thermal processing, or by thermal processing of the raw sludge alone. In the case of anaerobic digestion followed by thermal processing, roughly half of the available energy is captured as a medium BTU digester gas which is generally a more useable form of energy than waste heat for thermal processing. On the other hand, the digestion process adds to the total system costs. Therefore, trade-offs between having or not having a digestion process are difficult to assess except on a site specific basis. Treatment plants in the LA/OMA region currently employ anaerobic digestion and utilize the produced gas for power generation. It is likely that such facilities will continue to be operated. Therefore, the management alternative described in this chapter have all been assumed to employ anaerobic digestion.

From both practical and theoretical standpoints it is more efficient to dewater sludge prior to dehydration compared to feeding liquid sludge to the Carver-Greenfield process. It has therefore been assumed that mechanical dewatering would be an integral part of a sludge handling system involving the C-G process.

## DESCRIPTION OF THE C-G SYSTEM

A detailed diagram of a full-scale Carver-Greenfield process is presented in Figure 5.1. The four-effect system will be described below with the help of Figure 5.1. The system components were assembled by Dehydrotech Corporation and Foster-Wheeler Energy Corporation at the request of the LA/OMA Project. Mass and energy balances for the system will be discussed. Furthermore, comparisons will be made later in the chapter between systems employing the C-G process and other alternatives which do not employ dehydration.

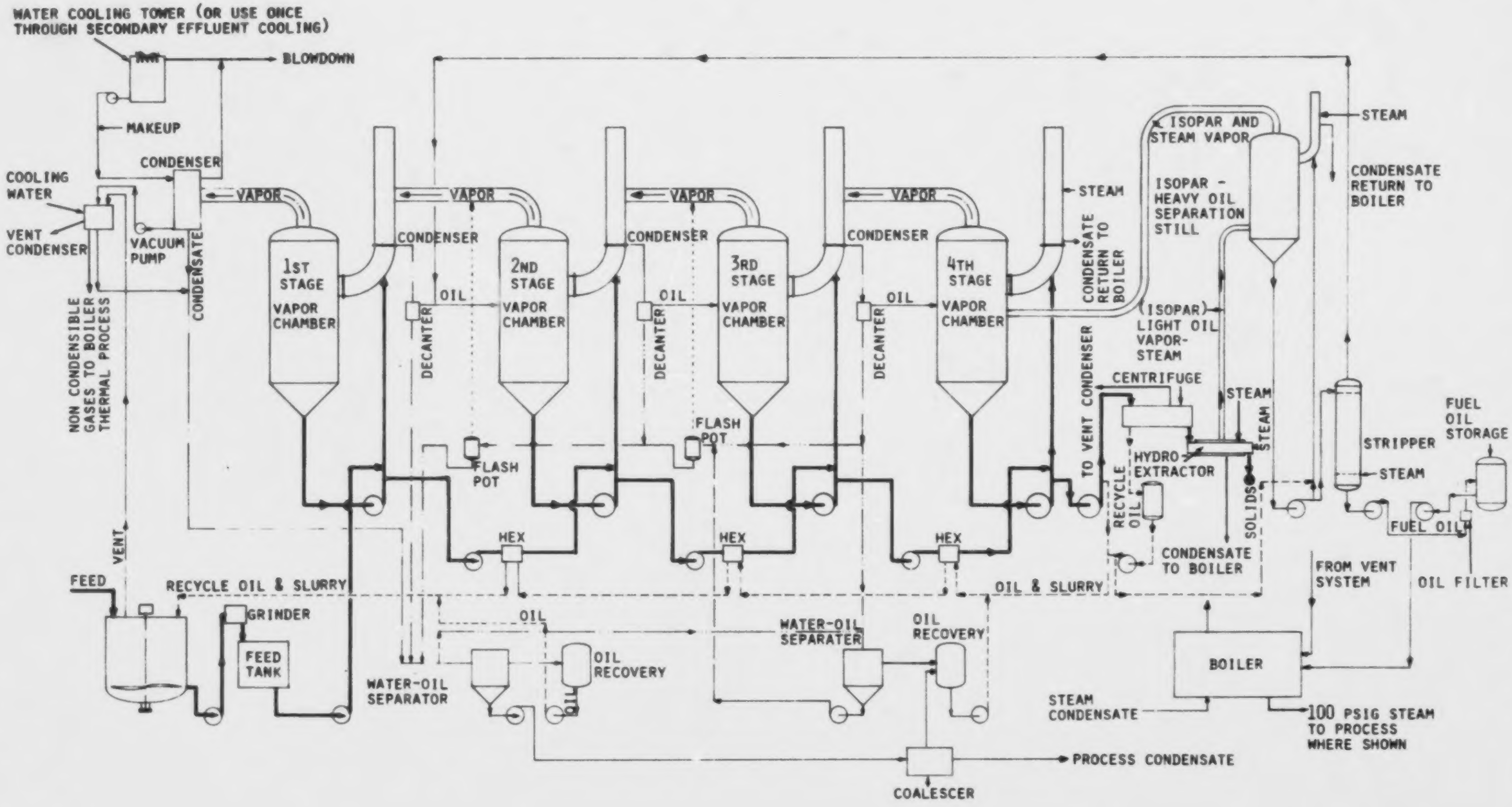
Referring to Figure 5.1, feed sludge enters the fluidizing tank where it is mixed with fluidizing oil. The latter is a mixture of recycle oil and dry sludge slurry obtained from the fourth stage evaporator. Capability for dry sludge slurry recycle, referred to as "addback" is included to improve the suspension qualities of the feed sludge and oil mixture. It also alleviates problems associated with gummy phase formation. A ratio of one part feed solids (dry basis) and one part of recycled dry slurry solids (oil free basis) has been recommended by the manufacturer.

The slurried mixture in the fluidizing tank is then pumped to a grinder having a ¼" size screen. Discharge from the grinder is fed by gravity to the evaporator feed tank fitted with an agitator to maintain a uniform slurry consistency.

Slurry is then pumped to the four effect evaporator system. In the evaporator system, temperature of the feed slurry increases from stage to stage whereas the water vapor flowing in opposite direction decreases in temperature from one effect to the next. The temperature difference between effects is necessary to allow heat transfer from vapor released in one effect to fluid contained in the next. The temperature difference is produced by maintaining progressively lower pressures in each effect and, therefore, progressively lower boiling points.

Slurry in the evaporator feed tank is pumped continuously to the first stage (fourth effect) where a constant fluid level is maintained. Flowrate to the first stage is modulated based on a level signal resulting from the head of the fluid mixture in the evaporator. Under normal operations, the flowrate is essentially constant. Each stage has a transfer pump and control system.

In order to recover heat energy from the hot recycle oil as well as from the fourth stage dry recycle slurry at about 255°F, three heat exchangers are included, as shown in Figure 5.1. As a result, temperature of the recycle solids-oil slurry is reduced from about 255°F to 145°F and thereby increasing the temperature of fluid being transferred between stages.


WATER COOLING TOWER (OR USE ONCE THROUGH SECONDARY EFFLUENT COOLING)
BLOWDOWN
MAKEUP
CONDENSER
COOLING WATER
VENT CONDENSER
VACUUM PUMP
CONDENSATE
NON CONDENSIBLE GASES TO BOILER THERMAL PROCESS
VENT
FEED
VAPOR
1ST STAGE VAPOR CHAMBER
2ND STAGE VAPOR CHAMBER
3RD STAGE VAPOR CHAMBER
4TH STAGE VAPOR CHAMBER
CONDENSER
OIL
DECANTER
FLASH POT
HEX
RECYCLE OIL & SLURRY
GRINDER
FEED TANK
WATER-OIL SEPARATOR
OIL
OIL RECOVERY
WATER-OIL SEPARATER
OIL RECOVERY
COALESCER
PROCESS CONDENSATE
OIL & SLURRY
STEAM
CONDENSATE RETURN TO BOILER
TO VENT CONDENSER
ISOPAR AND STEAM VAPOR
ISOPAR - HEAVY OIL SEPARATION STILL
(ISOPAR) LIGHT OIL VAPOR-STEAM
CENTRIFUGE
HYDRO-EXTRACTOR
RECYCLE OIL
CONDENSATE TO BOILER
SOLIDS
STEAM
CONDENSATE RETURN TO BOILER
STRIPPER
STEAM
FUEL OIL STORAGE
FUEL OIL
OIL FILTER
FROM VENT SYSTEM
BOILER
STEAM CONDENSATE
100 PSIG STEAM TO PROCESS WHERE SHOWN

REGIONAL WASTEWATER SOLIDS MANAGEMENT PROGRAM

**LA/OMA PROJECT**

LOS ANGELES/ORANGE COUNTY METROPOLITAN AREA

FLOW DIAGRAM OF A FULL SCALE FOUR-EFFECT CARVER-GREENFIELD PROCESS FIG. 5.1

It has been assumed that Isopar L, an aliphatic light oil, would be used as the fluidizing medium. At one atmosphere pressure the oil boils at 375 to 400°F decreasing to about 240 to 260°F at 0.10 atmospheres. The boiling point is considerably greater than that of water at all pressures which reduces the amount of oil vaporized along with water in the vapor chambers. However, the boiling point is sufficiently low that hydroextraction and steam stripping of Isopar is possible using steam of moderate temperature and pressure. Therefore, dry solids can be produced which contain only small amounts of the original fluidizing oil. An Isopar and steam mixture is azeotropic in that the mixture has a lower boiling point than either individual component. This is of some help in the hydroextraction and steam stripping operation. Since Isopar is a non-polar fluid it will tend to dissolve grease, fats and oils, and other non-polar materials contained in the sludge. Isopar can be steam stripped from most of these dissolved oils leaving a residue, hereinafter referred to as "heavy oil" which can be used as a fuel.

In most multi-effect systems, it is desirable to evaporate about equal quantities of water in each stage. However, heat recovery requirements often dictate varying water evaporation rates for different stages. In the case of the hydroextraction steam stripping technique, approximately equal quantities of water are evaporated in the first three stages and a lesser amount in the fourth stage whereas for the vacuum stripping and distillation, a more equal evaporation results in all four stages.

Condensate collected from the first two evaporation stages will contain distilled fluidizing oil in amounts up to about 40% or higher by weight. Vapor from the fourth stage will contain distilled oil in amounts up to about 100-150% of the water evaporated depending on the temperature of the slurry in the fourth stage.

Consider a single effect operating at 162°F. At this temperature the partial pressure of water and Isopar is about 5 psia and 0.60 psia, respectively. Since the moles of gas is proportional to the pressure the mole percentage of water to Isopar in the vapor would be about 8.3 to 1. However, actual weight percentage would be about 0.88 to 1 of water to Isopar. Even though the partial pressure of Isopar is significantly lower than water the higher molecular weight results in a large weight percentage being evaporated in each effect.

The water-oil condensate is decanted in small tanks and the oil is returned automatically to the stage from which it has been evaporated. An interface controller maintains a suitable oil level in each of these decantation tanks. Water is discharged to a larger oil-water separation tank where a more thorough separation of remaining oil from water is made. Since this water has been heated during the evaporation step, recovery of the heat is made by flashing to the higher vacuum of the preceeding stage. Two such flash units are employed in a four effect evaporator system. Trace quantities of oil remaining in the condensate are removed in a coalescer leaving a high quality water.

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A portion of the dry sludge-oil slurry (about one half) is recycled back through the heat exchangers to the fluidizing tank and the balance is centrifuged to produce dry solids containing about 40% oil. It is important to note that the dry sludge-oil slurry also contains the "heavy oil" which is approximately equal to 10% of the digested primary sludge feed (dry weight basis).

A portion of the recycled oil, approximately 25%, is pumped to the Heavy Oil Separation Still where 80% of the heavy oil is recovered. The balance of the oil is returned through heat exchangers to the fluidizing tank. Residual quantities of light oil leaving the Separation Still with the heavy oil is recovered through a vacuum stripper unit and the heavy oil is discharged from this unit. Heavy oil is then filtered and is thought to be suitable for use as a fuel oil.

If the heavy oil is not separated from the recycled Isopar the Isopar-heavy oil mixture would reach a saturation level. The heavy oil will then eventually show up in the centrifuged solids. Since only the Isopar, being a light oil, is distilled off in the hydroextraction process, heavier oils would tend to stay with the dried sludge. If dried sludge is to be thermally processed for energy recovery, heavy oil may not need to be separated from the sludge, and steam required to operate the C-G process could be supplied from the thermal processing system. In fact, keeping the heavy oils in the sludge is an advantage if the product is thermally processed. If separated, the heavy oil would have to be remixed with the solids or introduced separately into the pyrolysis unit. The latter can cause significant problems however due to the high BTU content of the heavy oil.

If energy recovery is not a part of the sludge handling scheme, then separation of the heavy oil becomes more important. The heavy oil could supplement the fuel requirement for the boiler operation which supplies steam to the C-G process in the latter case.

Centrifuged solids, containing light fluidizing oil and other dissolved heavy oila, enters a live bottom surge bin from where the solids are conveyed to a "Hydroextractor". The hydroextractor is a type of indirect steam dryer. Heat is supplied to the hydroextractor contents by indirect heat exchange with steam. Steam is also directly sparged through the product to steam distill the Isopar at a temperature of about 250°F to 325°F at atmospheric pressure. An alternate technique is to distill the Isopar without any direct steam input at a vacuum of 15 inch absolute. Energy required to raise the temperature of the product to 325°F and provide the heat of vaporization of the oil is obtained by supplying steam to the internal hollow mixing conveyor flights and the steam jacket of the hydroextractor unit. Vapors of the hydroextraction unit enter the light oil-heavy oil Separation Still to separate the oil fractions. Isopar and steam are then returned to the 4th stage vapor chamber to recover the heat value.

Dry product is withdrawn from the system after hydroextraction. Process condensate is recycled back to the treatment plant. All vent gases after condensation for oil recovery are combusted in the boiler or thermal reactor.

Mass and energy balances for a four-effect C-G installation processing 20 ton of dry sludge were estimated by the manufacturer and by staff of the Los Angeles County Sanitation Districts. For a four-effect C-G process, it was estimated that 0.44 pound steam would be required to evaporate one pound of water from sludge. The estimate made by LACSD staff is shown in Appendix 2.

## SYSTEM COMPARISON

Complete systems of sludge handling employing the Carver-Greenfield process could include at least the following three process trains:

- Mechanical Dewatering of Combined Sludge - Dehydration - Trucking to Landfill
- Mechanical Dewatering of Combined Sludge - Dehydration - Sale of Dried Product
- Mechanical Dewatering of Combined Sludge - Dehydration - Energy Recovery - Ash Trucked to Landfill

For the purpose of the present discussion, the City of Los Angeles' Hyperion Treatment Plant was used as an example for case studies. It was assumed that all sludge would be anaerobically digested and dewatered to 25 percent T.S. Since dewatering is common of all the above process trains, the following analyses considered only the processes downstream of mechanical dewatering. The above mentioned process trains were included in the comparison with the exception of sale of Carver-Greenfield dried product. Other "typical" sludge management concepts were also included for comparative purposes. A brief description of all systems considered is presented in Table 5.1.

TABLE 5.1

SYSTEM DESCRIPTIONS

| System No. | Descriptions |
|---|---|
| 1 | Sludge Cake Trucked to Landfill |
| 2 | Sludge Cake Dehydrated Prior to Trucking to Landfill |
| 3 | Sludge Cake Dehydrated, Thermally Processed for Energy Recovery and Residue Trucked to Landfill |

| System No. | Description |
|---|---|
| 4 | Sludge Cake Thermally Processed with Fuel Oil as Supplemental Fuel for Autogenous Combustion |
| 5 | Sludge Cake Thermally Processed with Refuse Derived Fuel (RDF) as Supplemental Fuel for Autogenous Combustion. Sludge (dry) to RDF (as received) Ratio of 1:1.4 |
| 6 | Same as System 5 except sludge (dry) to RDF (as received) Ratio of 1.6. |
| 7 | Thermal Processing - C-G dried sludge blended with RDF |

MAJOR ASSUMPTIONS USED IN ANALYSIS

- For all alternatives, raw primary sludge is blended with thickened waste activated sludge, anaerobically digested, and mechanically dewatered.
- Sludge production after dewatering is 370 dtpd at 25 percent total solids and 65 percent volatile fraction.
- All capital costs for the unit processes are adjusted to an Engineering News Record (ENR) Index of 3031. Equipment and buildings are amortized at 6 5/8percent for 20 years, with the exception of trucks which are amortized for 5 years. All operation and maintenance costs are based on a Consumer Price Index (CPI) of 183.
- 30-mile one-way haul distance to landfill for all residues.
- All sludge and refuse processing on existing treatment plant sites.
- Production of Carver-Greenfield product at 344 dtpd containing 217 dtpd VS and 23 tpd of extracted heavy oil blended with CG dried solids prior to thermal processing.
- Thermal processing for all systems conducted in a multiple hearth furnace operated under a starved air pyrolysis mode.
- Refuse processing includes shredding, magnetic separation, and air classification to produce an RDF supplementary fuel.

- Credits and other cost factors are as shown in Tables 5.2 through 5.8.
- Design criteria for each unit process are as presented in Appendix III.

Data used for comparative analysis were derived from information provided in the analyses of the LA/OMA Sludge Management Alternatives (15). In general, the cost estimates correspond to "order of magnitude" estimates as defined by the American Association of Cost Engineers. Relative costs between the various permutations, rather than actual costs, are therefore of primary concern. To the extent possible major assumptions used in developing the cost and the energy balances have been documented and are included in Tables 5.1 to 5.7 and Appendix III. However, for further detailed assumptions the reader should consult reference (15).

Tables 5.2 through 5.7 show the various cost factors and some of the other important parameters for each of the six systems. Only unit processes downstream of mechanical dewatering are included in the analysis. For example, figures shown for electricity produced do not include contributions from digester gas. Table 5.8 and Figure 5.2 summarize the cost and energy data developed for the six systems.

## DEWATERING AND LANDFILLING

Under this concept two alternative systems were considered as shown in Figure 5.3. One alternative involved direct trucking of dewatered sludge cake to a landfill (System 1 - Table 5.2). The other alternative assumed that dewatered sludge would be dehydrated using the C-G process prior to landfilling (System 2 - Table 5.3). The same haul distance was assumed in both cases. Data presented in Tables 5.2 and 5.3 indicate that trucking sludge cake (System 1) is substantially lower in cost than drying with the C-G process before trucking (System 2). System 1 was lower in both capital, and operation and maintenance (O & M) costs.

Dehydrating the sludge prior to hauling would save about 960 gal/day of diesel fuel for the trucking operation. However, this would be more than compensated by about 4160 gal/day of fuel oil required to operate the CG boiler in addition to the energy value of 983 gal/day of Isopar L. Under the conditions assumed, dehydration before landfilling would be more energy intensive unless an alternate source of energy were available, for example, flared or unused digester gas. Diesel fuel saved would then offset the Isopar needed for dehydration. In this same regard the cost of fuel oil for the C-G unit is a major O & M expense. If currently unused digester gas were substituted for fuel oil the cost difference would reduce to only about $3/ton. Therefore, under certain combinations of conditions -- such as an inexpensive energy source (if such can be found today), very long haul distances, and high landfill disposal fees -- prior dehydration could be cost competitive.

## TABLE 5.2

## COST AND ENERGY FIGURES FOR SLUDGE MANAGEMENT SYSTEM

System No. 1
Description: Trucking to Landfill

| | Truck Transportation & Disposal |
|---|---|
| **Sizing** | |
| Land Area (acres) | |
| No. of Employee Positions/24 hrs. | 33 |
| Isopar L used (gal/day) | |
| RDF used (tpd) | |
| Added $O_2$ for recycle (tpd) | |
| **Energy** | |
| Elec. used (KWH/day) | |
| Elec. prod. (KWH/day) | |
| Fuel Oil used (gal/day) | |
| Diesel fuel used (gal/day) | 1320 |
| Elec. for recycle $O_2$ (KWH/day) | |
| **Capital** | |
| Process ($1000) | 2376 |
| Building ($1000) | |
| Subtotal Annual Capital ($1000/yr) | 574 |
| **O & M** ($1000/yr) | |
| Labor ($43,300/position-yr) | 1430 |
| Diesel Fuel ($.40/gal) | 193 |
| Fuel Oil ($.40/gal) | |
| Electricity ($.03/KWH) | |
| Isopar L ($.70/gal) | |
| Maintenance | 489 |
| Recycle $O_2$ | |
| Dump Charge ($2.50/ton) | 1345 |
| Subtotal O & M | 3457 |
| **Credits** ($1000/yr) | |
| Electricity Produced ($.03/KWH) | |
| Refuse Tipping Fee ($10/ton) | |
| Refuse Resale | |
| Subtotal Credits | 0 |
| **Total Annual Cost** ($1000/yr) | 4031 |
| ($/ton)* | 29.90 |

* On the basis of 370 dtpd dewatered cake

## TABLE 5.3

## COST AND ENERGY FIGURES FOR SLUDGE MANAGEMENT SYSTEM

System No. 2
Description: C-G and Trucking

| | C-G | Trucking & Disp. | Total |
|---|---|---|---|
| **Sizing** | | | |
| Land Area (acres) | 0.9 | | 0.9 |
| No. of Employee Positions/24 hrs. | 8 | 9 | 17 |
| Isopar L used (gal/day) | 983 | | 983 |
| RDF used (tpd) | | | |
| Added $O_2$ for recycle (tpd) | 1.7 | | 1.7 |
| **Energy** | | | |
| Elec. used (KWH/day) | 36500 | | 36500 |
| Elec. prod. (KWH/day) | | | |
| Fuel Oil used (gal/day) | 4160 | | 4160 |
| Diesel fuel used (gal/day) | | 360 | 360 |
| Elec. for recycle $O_2$ (KWH/day) | 540 | | 540 |
| **Capital** | | | |
| Process ($1000) | 23100 | 700 | 23800 |
| Building ($1000) | 2000 | | 2000 |
| Subtotal Annual Capital ($1000/yr) | 2300 | 170 | 2470 |
| **O & M** ($1000/yr) | | | |
| Labor ($43,300/position-yr) | 350 | 380 | 730 |
| Diesel Fuel ($.40/gal) | | 50 | 50 |
| Fuel Oil ($.40/gal) | 610 | | 610 |
| Electricity ($.03/KWH) | 400 | | 400 |
| Isopar L ($.70/gal) | 250 | | 250 |
| Maintenance | 70 | 130 | 200 |
| Recycle $O_2$ | 10 | | 10 |
| Dump Charge ($2.50/ton) | | 350 | 350 |
| Subtotal O & M | 1690 | 910 | 2600 |
| **Credits** ($1000/yr) | | | |
| Electricity Produced ($.03/KWH) | | | |
| Refuse Tipping Fee ($10/ton) | | | |
| Refuse Resale | | | |
| Subtotal Credits | 0 | 0 | 0 |
| **Total Annual Cost** ($1000/yr) | 3990 | 1080 | 5070 |
| ($/ton)* | 29.50 | 8.00 | 37.50 |

* On the basis of 370 dtpd dewatered cake

## TABLE 5.4

## COST AND ENERGY FIGURES FOR SLUDGE MANAGEMENT SYSTEM

System No. 3
Description: C-G with Energy Recovery

| | C-G | DFTP | Trucking & Disposal | Total |
|---|---|---|---|---|
| Sizing | | | | |
| Land Area (acres) | 0.9 | 0.6 | | 1.5 |
| No. of Employee Positions/24 hrs. | 8 | 5 | 3 | 16 |
| Isopar L used (gal/day) | 980 | | | 980 |
| RDF used (tpd) | | | | |
| Added $O_2$ for recycle (tpd) | 1.7 | | | 1.7 |
| Energy | | | | |
| Elec. used (KWH/day) | 36500 | 33100 | | 69600 |
| Elec. prod. (KWH/day) | | 181900 | | 181900 |
| Fuel Oil used (gal/day) | | | | 0 |
| Diesel fuel used (gal/day) | | | 130 | 130 |
| Elec. for recycle $O_2$ (KWH/day) | 407 | | | 407 |
| Capital | | | | |
| Process ($1000) | 23100 | 17100 | 200 | 43100 |
| Building ($1000) | 2000 | 700 | (Total) | (Total) |
| Subtotal Annual Capital ($1000/yr) | 2300 | 1630 | 50 | 3980 |
| O & M ($1000/yr) | | | | |
| Labor ($43,300/position-yr) | 350 | 220 | 130 | 700 |
| Diesel Fuel ($.40/gal) | | | 20 | 20 |
| Fuel Oil ($.40/gal) | | | | |
| Electricity ($.03/KWH) | 400 | 320 | | 720 |
| Isopar L ($.70/gal) | 250 | | | 250 |
| Maintenance | 70 | 50 | 40 | 160 |
| Recycle $O_2$ | 10 | | | 10 |
| Dump Charge ($2.50/ton) | | | 120 | 120 |
| Subtotal O & M | 1080 | 590 | 310 | 1980 |
| Credits ($1000/yr) | | | | |
| Electricity Produced ($.03/KWH) | | -1990 | | -1990 |
| Refuse Tipping Fee ($10/ton) | | | | |
| Refuse Resale | | | | |
| Subtotal Credits | | -1990 | | -1990 |
| Total Annual Cost ($1000/yr) | 3380 | 230 | 360 | 3970 |
| ($/ton)* | 25.00 | 1.70 | 2.70 | 29.40 |

* On the basis of 370 dtpd dewatered cake

## TABLE 5.5

## COST AND ENERGY FIGURES FOR SLUDGE MANAGEMENT SYSTEM

System No. 4
Description: Thermal Processing - Fuel Oil Used as Supplemental Fuel

| | DFTP | Trucking & Disposal | Total |
|---|---|---|---|
| Sizing | | | |
| Land Area (acres) | 1.4 | | 1.4 |
| No. of Employee Positions/24 hrs. | 10 | 3 | 13 |
| Isopar L used (gal/day) | | | |
| RDF used (tpd) | | | |
| Added $O_2$ for recycle (tpd) | | | |
| Energy | | | |
| Elec. used (KWH/day) | 110600 | | 110600 |
| Elec. prod. (KWH/day) | 162300 | | 162300 |
| Fuel Oil used (gal/day) | 9020 | | 9020 |
| Diesel fuel used (gal/day) | | 130 | 130 |
| Elec. for recycle $O_2$ (KWH/day) | | | |
| Capital | | | |
| Process ($1000) | 46400 | 200 | 46600 |
| Building ($1000) | 2100 | | 2100 |
| Subtotal Annual Capital ($1000/yr) | 4440 | 50 | 4490 |
| O & M ($1000/yr) | | | |
| Labor ($43,300/position-yr) | 430 | 130 | 560 |
| Diesel Fuel ($.40/gal) | | 20 | 20 |
| Fuel Oil ($.40/gal) | 1320 | | 1320 |
| Electricity ($.03/KWH) | 1210 | | 1210 |
| Isopar L ($.70/gal) | | | |
| Maintenance | 130 | 40 | 170 |
| Recycle $O_2$ | | | |
| Dump Charge ($2.50/ton) | | 120 | 120 |
| Subtotal O & M | 3090 | 310 | 3400 |
| Credits ($1000/yr) | | | |
| Electricity Produced ($.03/KWH) | -1780 | | -1780 |
| Refuse Tipping Fee ($10/ton) | | | |
| Refuse Resale | | | |
| Subtotal Credits | -1780 | | -1780 |
| Total Annual Cost ($1000/yr) | 5750 | 360 | 6110 |
| ($/ton)* | 42.60 | 2.70 | 45.30 |

* On the basis of 370 dtpd dewatered cake

## TABLE 5.6

## COST AND ENERGY FIGURES FOR SLUDGE MANAGEMENT SYSTEM

System No. 5
Description: Thermal Processing - RDF Used as Supplemental Fuel (Sludge:RDF = 1:1.14) Autogenous Case

| | RDF Processing | DFTP | Trucking & Disposal | Total |
|---|---|---|---|---|
| Sizing | | | | |
| Land Area (acres) | 4 | 2 | | 6 |
| No. of Employee Positions/24 hrs | 19 | 22 | 9 | 50 |
| Isopar L used (gal/day) | | | | |
| RDF used (tpd) | 420 | | | 420 |
| Added $O_2$ for recycle (tpd) | | | | |
| Energy | | | | |
| Elec. used (KWH/day) | 15000 | 142000 | | 157000 |
| Elec. prod.(KWH/day) | 0 | 244000 | | 244000 |
| Fuel Oil used (gal/day) | | | | 0 |
| Diesel fuel used (gal/day) | 370 | | 360 | 730 |
| Elec. for recycle $O_2$ (KWH/day) | | | | 0 |
| Capital | | | | |
| Process ($1000) | 6700 | 62800 | | |
| Building ($1000) | 600 | 2800 | 700 | 73600 |
| Subtotal Annual Capital($1000/yr) | 680 | 6010 | 160 | 6850 |
| O & M ($1000/yr) | | | | |
| Labor ($43,300/position-yr) | 590 | 950 | 310 | 1850 |
| Diesel Fuel ($.40/gal) | 40 | | 60 | 100 |
| Fuel Oil ($.40/gal) | | | | 0 |
| Electricity ($.03/KWH) | 120 | 1560 | | 1680 |
| Isopar L ($.70/gal) | | | | 0 |
| Maintenance | 270 | 180 | 110 | 560 |
| Recycle $O_2$ | | | | 0 |
| Dump Charge ($2.50/ton) | | | 260 | 260 |
| Subtotal O & M | 1020 | 2690 | 740 | 4450 |
| Credits ($1000/yr) | | | | |
| Electricity Produced ($.03/KWH) | | -2670 | | -2670 |
| Refuse Tipping Fee ($10/ton) | -1950 | | | -1950 |
| Refuse Resale | -440 | | | -440 |
| Subtotal Credits | -2390 | -2670 | | -5060 |
| Total Annual Cost ($1000/yr) | -690 | 6030 | 900 | 6320 |
| ($/ton)* | -5.10 | 44.70 | 6.70 | 46.30 |

* On the basis of 370 dtpd dewatered cake

## TABLE 5.7

## COST AND ENERGY FIGURES FOR SLUDGE MANAGEMENT SYSTEM

System No. 6
Description: Thermal Processing-RDF Used as Supplemental Fuel (Sludge: RDF = 1:1.6

| | RDF Processing | DFTP | Trucking & Disposal | Total |
|---|---|---|---|---|
| **Sizing** | | | | |
| Land Area (acres) | 6 | 2 | - | 8 |
| No. of Employee Positions/24 hrs. | 27 | 22 | 12 | 61 |
| Isopar L used (gal/day) | | | | 0 |
| RDF used (tpd) | 590 | | | 590 |
| Added $O_2$ for recycle (tpd) | | | | 0 |
| **Energy** | | | | |
| Elec. used (KWH/day) | 21000 | 155000 | | 176000 |
| Elec. prod. (KWH/day) | | 334000 | | 334000 |
| Fuel Oil used (gal/day) | | | | 0 |
| Diesel fuel used (gal/day) | 520 | | 440 | 960 |
| Elec. for recycle $0_2$ (KWH/day) | | | | 0 |
| **Capital** | | | | |
| Process ($1000) | 8100 | 63900 | 900 | 75700 |
| Building ($1000) | 900 | 2800 | (Total) | (Total) |
| Subtotal Annual Capital ($1000/yr) | 820 | 6110 | 210 | 7140 |
| **O & M** ($1000/yr) | | | | |
| Labor ($43,300/position-yr) | 840 | 950 | 520 | 2310 |
| Diesel Fuel ($.40/gal) | 60 | | 60 | 120 |
| Fuel Oil ($.40/gal) | | | | 0 |
| Electricity ($.03/KWH) | 160 | 1700 | | 1860 |
| Isopar L ($.70/gal) | | | | 0 |
| Maintenance | 320 | 180 | 140 | 640 |
| Recycle $0_2$ | | | | 0 |
| Dump Charge ($2.50/ton) | | | 320 | 320 |
| Subtotal O & M | 1380 | 2830 | 1040 | 5250 |
| **Credits** ($1000/yr) | | | | |
| Electricity Produced ($.03/KWH) | | -3660 | | -3660 |
| Refuse Tipping Fee ($10/ton) | -2730 | | | -2730 |
| Refuse Resale | -620 | | | -620 |
| Subtotal Credits | -3350 | -3660 | 0 | -7010 |
| **Total Annual Cost** ($1000/yr) | -1150 | 5280 | 1250 | 5380 |
| ($/ton)* | -8.50 | 39.10 | 9.30 | 39.90 |

* On the basis of 370 dtpd dewatered cake

## TABLE 5.8

## COST AND ENERGY FIGURES FOR SLUDGE MANAGEMENT SYSTEM

System No. 7
Description: Thermal Processing - C-G Dried Sludge Blended with RDF

| | C-G | RDF Processing | DFTP | Trucking & Disposal | Total |
|---|---|---|---|---|---|
| Sizing | | | | | |
| Land Area (acres) | 0.9 | 4 | 1.5 | | 6.4 |
| No. Of Employee Positions/24 hrs | 8 | 19 | 11 | 9 | 47 |
| Isopar L used (gal/day) | 980 | -- | -- | -- | 980 |
| RDF used (tpd) | -- | 750+ | 589 | -- | 590 |
| Added $O_2$ for Recycle (tpd) | 1.7 | | | | |
| Energy | | | | | |
| Elec. used (KWH/Day) | 36500 | 15000 | 61000 | | 112500 |
| Elec. prod. (KWH/Day) | | | 483000 | | 483000 |
| Fuel Oil used (gal/day) | | | | | |
| Diesel fuel used (gal/day) | | 370 | | 360 | 730 |
| Elec. for recycle $O_2$ (KWH/day) | 407 | | | | 407 |
| Capital | | | | | |
| Process ($1000) | 23100 | 6723 | 35750 | 650 | 70253 |
| Building ($1000) | 2000 | 640 | 1390 | (Total) | (Total) |
| Subtotal Annual Capital ($1000/yr) | 2300 | 675 | 3400 | 157 | 6532 |
| O & M | | | | | |
| Labor ($43,000/position-yr) | 350 | 590 | 480 | 390 | 1810 |
| Diesel Fuel ($.40/gal) | | 39 | | 53 | 92 |
| Fuel Oil ($.40/gal) | | | | | |
| Electricity ($.03/KWH) | 400 | 117 | 660 | | 1177 |
| Isopar L ($.70/gal) | 250 | | | | 250 |
| Maintenance | 70 | 265 | 100 | 113 | 548 |
| Recycle $O_2$ | 10 | | | 261 | 271 |
| Dump Charge ($2.50/ton) | | | | | |
| Subtotal O & M | 1080 | 1011 | 1240 | 817 | 4148 |
| Credits ($1000/yr) | | | | | |
| Elec. Produced ($.03/KWH) | | | -5290 | | -5290 |
| Refuse Tipping Fee ($10/Ton) | | -1950 | | | -1950 |
| Refuse Resale | | -443 | | | -443 |
| Subtotal Credits | | -2393 | -5290 | | -7683 |
| Total Annual Cost ($1000/yr) | 3380 | -707 | -650 | 974 | 2997 |
| ($/ton)* | 25.00 | -5.20 | -4.80 | 7.20 | 22.20 |

* On the basis of 370 dtpd dewatered cake
+ Refuse input

## TABLE 5.9

## SUMMARY OF SYSTEM COSTS

| System | Total Net Cost ($/dry ton) | Cost in $1000/year Ammortized | Cost in $1000/year O & M | Cost in $1000/year Credit |
|---|---|---|---|---|
| 1 | 29.90 | 574 | 3457 | 0 |
| 2 | 37.50 | 2470 | 2600 | 0 |
| 3 | 29.40 | 3980 | 1980 | -1990 |
| 4 | 45.30 | 4490 | 3400 | -1780 |
| 5 | 46.30 | 6850 | 4450 | -5060 |
| 6 | 39.90 | 7140 | 5250 | -7010 |
| 7 | 22.20 | 6530 | 4150 | -7690 |

## TABLE 5.10

## SYSTEM ENERGY BALANCE

| SYSTEM | ENERGY USE |  |  |  | ENERGY PRODUCED |
|---|---|---|---|---|---|
|  | Electricity | Fuel Oil |  | Total |  |
|  | $10^3$KWH/Day | Gal/Day | $10^3$KWH/Day[a/] | $10^3$KWH/Day | $10^3$KWH/Day |
| 1 | 0 | 1320 | 17.8 | 17.8 | 0 |
| 2 | 36.5 | 4520 | 60.9 | 97.4 | 0 |
| 3 | 69.6 | 130 | 1.8 | 71.4 | 182 |
| 4 | 110.6 | 9150 | 123.3 | 233.9 | 162 |
| 5 | 157 | 730 | 9.8 | 166.8 | 244 |
| 6 | 176 | 960 | 13 | 189 | 334 |
| 7 | 112 | 730 | 9.8 | 121.8 | 483 |

a/ Assumes 140,000 BTU/GAL and 10,387 BTU/KWH

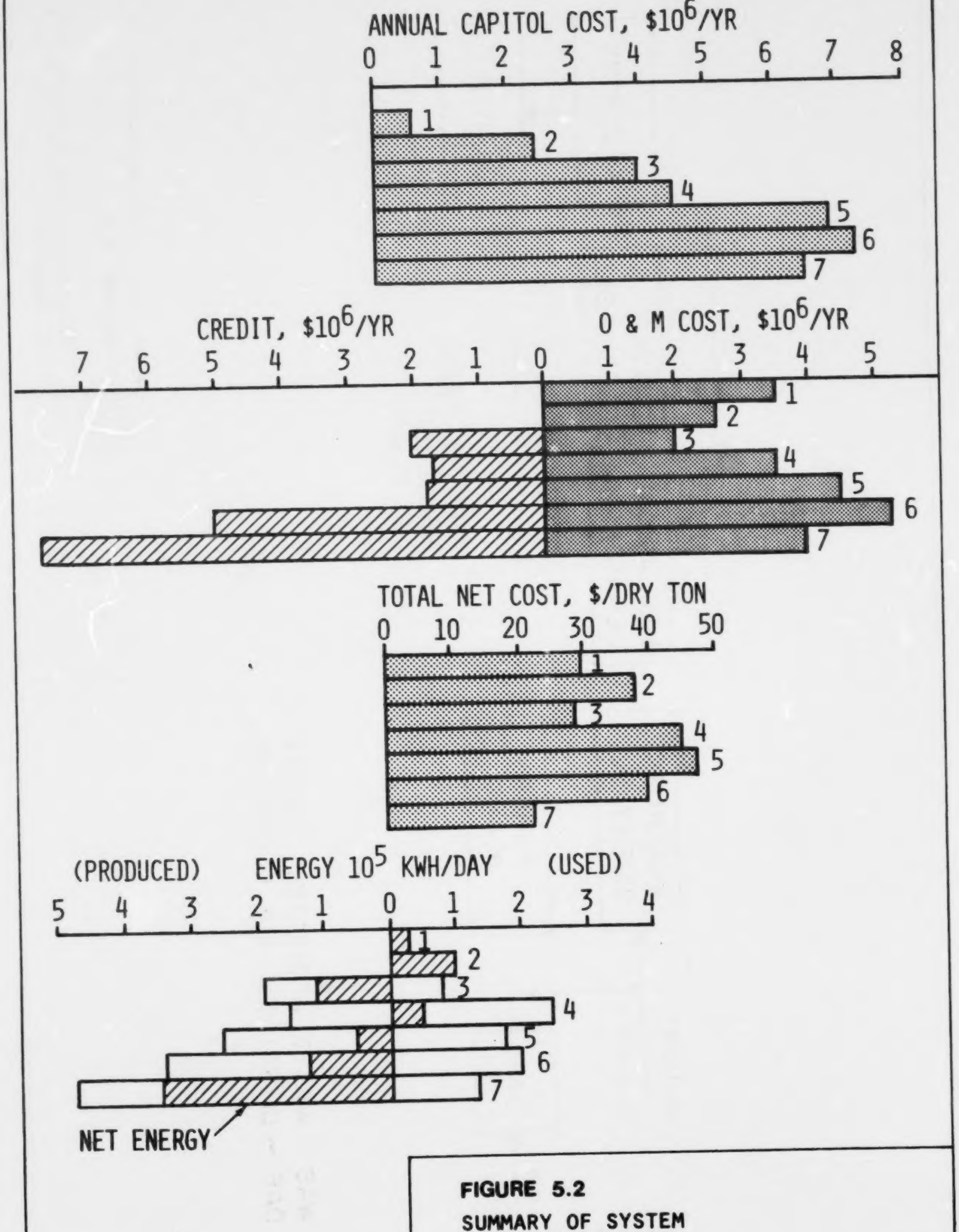
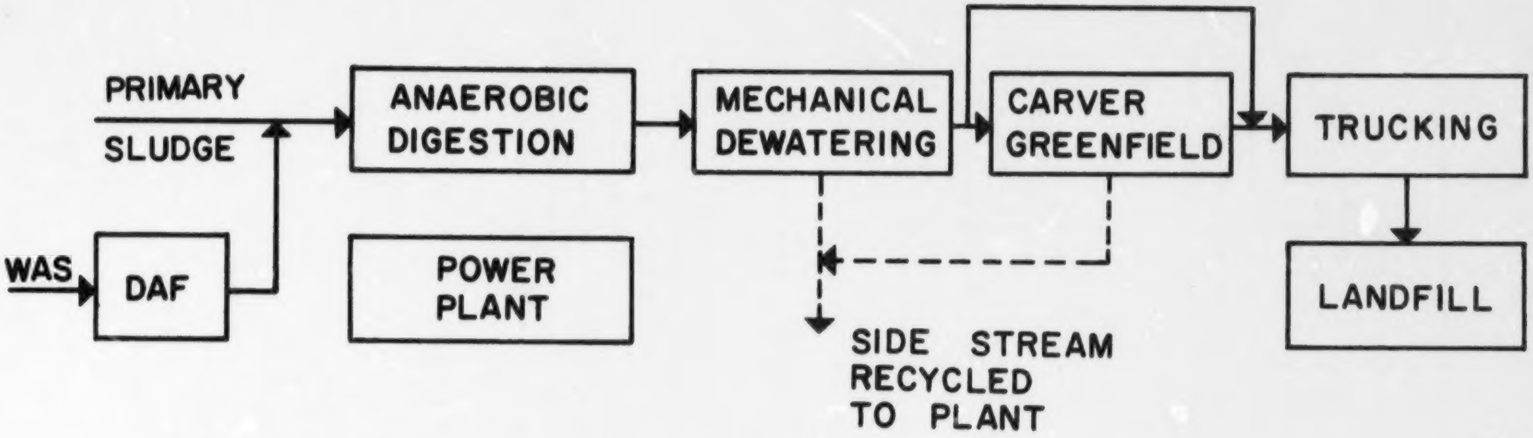


**FIGURE 5.2**

SUMMARY OF SYSTEM
COSTS AND ENERGY


PRIMARY
SLUDGE
WAS
DAF
ANAEROBIC DIGESTION
POWER PLANT
MECHANICAL DEWATERING
CARVER GREENFIELD
TRUCKING
LANDFILL
SIDE STREAM RECYCLED TO PLANT

WAS — WASTE ACTIVATED SLUDGE
DAF — DISSOLVED AIR FLOTATION

FIGURE 5.3

DEWATERING AND LANDFILLING OF SLUDGE

It should, however, be pointed out that direct hauling to a landfill without a dehydration process, involved the transport of about four times more total weight of sludge. As this involves a proportionately higher number of trucks and higher dump fees, the system would be very sensitive to rises in the cost of trucking and dump fees. Furthermore, dried sludge from the C-G process could be placed in a Class II landfill. On the other hand, all of the wet sludge cake from the other system may not be acceptable in a single Class II landfill due to the moisture release from the sludge compared with the absorption capacity of the available refuse. However, the fluffy nature of the dried sludge from C-G process could give rise to material handling problems during trucking and landfilling. If the dry sludge produced by the C-G process could be sold instead of landfilling, system economics would improve somewhat.

## ENERGY RECOVERY FROM SLUDGE USING EITHER CARVER-GREENFIELD PROCESS OR SUPPLEMENTAL FUEL

Under this mode of sludge management concept sludge would be thermally processed with energy recovery in the form of steam and electricity. Figure 5.4 shows two alternative process trains to achieve this objective. Under one alternative, sludge cake would be dried to 95 percent solids by using the Carver-Greenfield process before being fed into the thermal reactor. In the other alternative, instead of drying the sludge cake, supplemental fuel is added in the burning step to achieve autogenous combustion. Supplemental fuel could be either fuel oil or, most likely, refuse derived fuel (RDF). Both forms of supplemental fuel were considered by the LA/OMA Project during analyses of the various permutations of the Sludge Management Alternatives (15). In the present discussion, the C-G process will be compared against use of both fuel oil and RDF, respectively.

The sludge drying alternative considers dewatering of digested combined sludges followed by drying and thermal processing (System 3 - Table 5.4). Under alternatives using supplemental fuels, three systems were considered. One system involved use of fuel oil to achieve autogenous combustion (System 4 - Table 5.5). In the other two systems RDF was added in two different sludge (dry) to RDF (as received) ratios by weight; 1.14 to achieve autogenous combustion (System 5 - Table 5.6), and 1.6 to generate more electricity (System 6 - Table 5.7).

In the present example, multiple hearth furnaces were assumed for the direct fired thermal processing (DFTP) step. A schematic of the thermal processing-energy recovery concept is shown in Figure 5.5. Feed to the reactor would be partially oxidized and gasified using less than stoichiometric amounts of air. The resultant fuel gas would be combusted in an afterburner. Included in the process are heat recovery via a waste heat boiler and electrical power generation in a steam turbine. Also included are emission control to reduce air pollution using high energy venturi scrubbers followed in series by multiple tray scrubbers. In all cases,

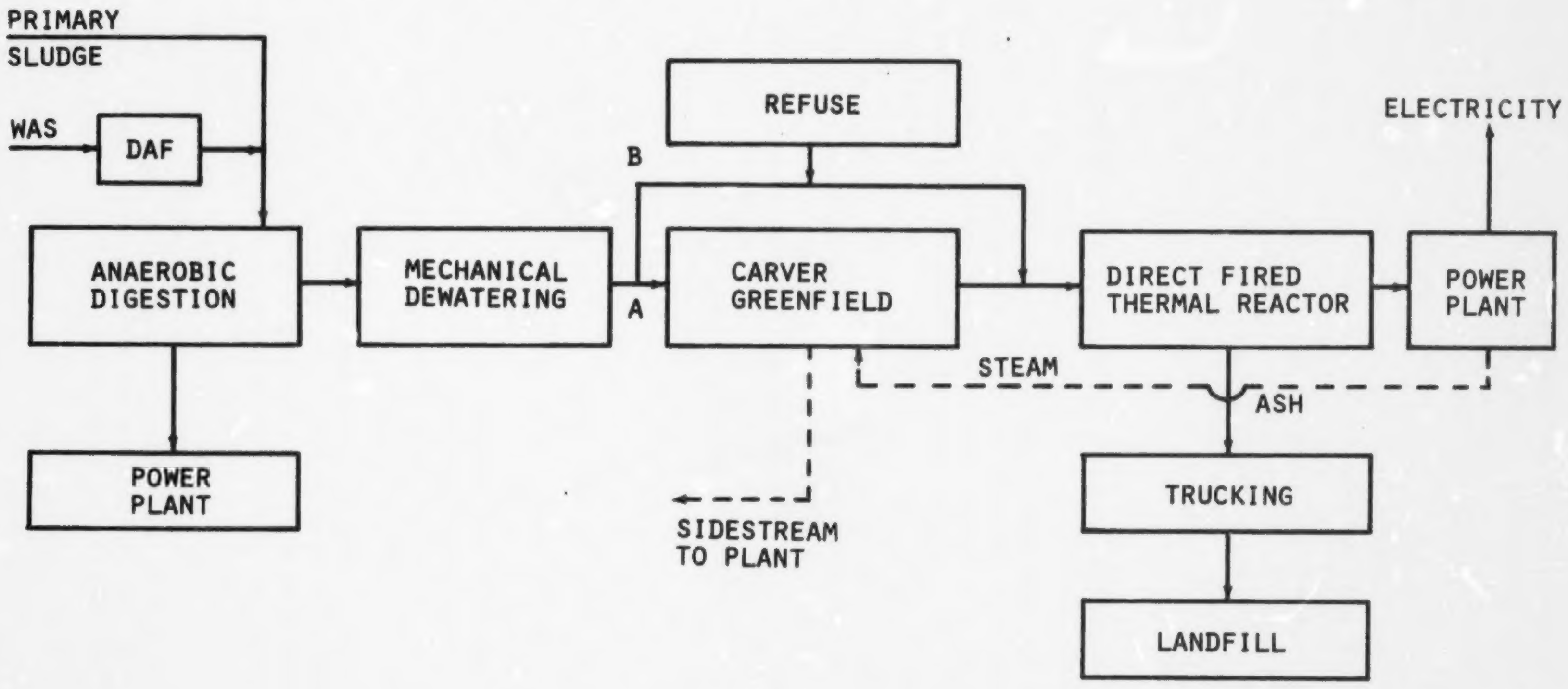

PRIMARY
SLUDGE
WAS
DAF
ANAEROBIC DIGESTION
POWER PLANT
MECHANICAL DEWATERING
A
B
REFUSE
CARVER GREENFIELD
SIDESTREAM TO PLANT
STEAM
DIRECT FIRED THERMAL REACTOR
POWER PLANT
ELECTRICITY
ASH
TRUCKING
LANDFILL

**FIGURE 5.4**
ENERGY RECOVERY FROM SLUDGE WITH EITHER CARVER-GREENFIELD PROCESS OR REFUSE ADDICTION

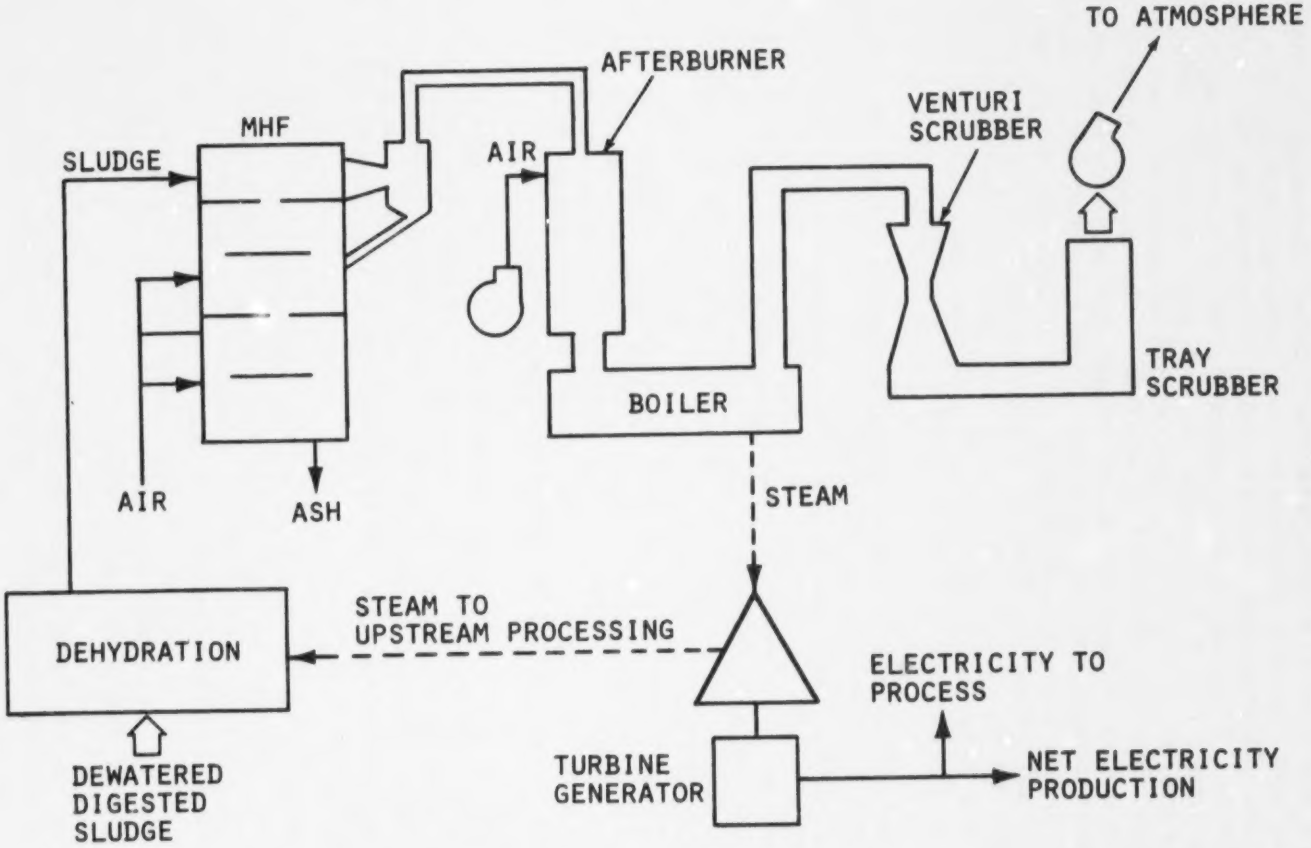

TO ATMOSPHERE
AFTERBURNER
VENTURI SCRUBBER
MHF
SLUDGE
AIR
AIR
ASH
BOILER
TRAY SCRUBBER
STEAM
STEAM TO UPSTREAM PROCESSING
DEHYDRATION
DEWATERED DIGESTED SLUDGE
ELECTRICITY TO PROCESS
TURBINE GENERATOR
NET ELECTRICITY PRODUCTION

**FIGURE 5.5**

THERMAL PROCESSING-ENERGY RECOVERY CONCEPT

LA/OMA PROJECT

afterburner outlet temperature was raised to a minimum of 1400°F. Pertinent design criteria used in the engineering and economic analyses of those alternatives involving use of direct fired thermal processing are listed in Appendix III.

Tables 5.4 through 5.8 indicate that under the conditions assumed, C-G drying and thermal processing of combined sludge for energy recovery has substantially lower cost compared to all other alternatives. Although the systems using refuse as supplemental fuel have lower net O & M costs (O & M less credit), they also have substantially higher capital costs because of the increased size of the thermal processing system. The higher number of furnaces and the refuse handling needs are also reflected in higher labor costs for the systems involving refuse. Although Systems 5 and 6 received a credit of $10 per ton of refuse used in the thermal process, System 3 employing C-G process would be less expensive by about $10 to $17 per ton of sludge (dry) processed. For the Hyperion Treatment Plant this would result in an annual cost saving of 1.4 to 2.3 million dollars.

However, both Systems 5 and 6 generate significant amount of revenue from sale of surplus electricity. In the present analysis electricity cost has been assumed at 3¢/KWH. With higher cost of electricity, which is highly probable in the near future, the cost differences between System 3 and Systems 5 and 6 will decrease. However, analysis of the data in Tables 5.4 and 5.6 indicates that electricity would have to increase to about 12¢/KWH before System 6 became equal in annual cost to System 3. All other variables held constant. On the other hand, costs for Systems 5 and 6 are very sensitive to the assumed refuse tipping fee. If the tipping fee was increased to about $15/ton as received refuse, instead of the assumed $10/ton, total annual costs for Systems 6 and 7 would be about equal to that for System 3.

Both Systems 5 and 6 are net revenue producing. In otherwords, they generate higher credits than their respective O & M costs. This might seem attractive to public operating agencies since the O & M cost has to be borne from their annual operating budget. A major share of the capital cost comes from the federal and state funding agencies.

It would be well to mention at this point that the assumed electrical credit of 3¢/KWH is not meant to imply that a treatment agency would actually be selling electricity at that price. In all likelihood the electricity would be used within the wastewater treatment plant in place of ther electrical supplies. The credit implies that the electricity has an economic value and not that actual revenues would be generated.

It should be noted that significant quantities of refuse are required to achieve autogenous combustion under conditions established for operation of the multiple hearth furnace. In general, about 1.14 wet tons of refuse derived fuel (RDF) are required for every dry ton of sludge. At Hyperion, for processing 370 dry tons/per day of digested sludge cake, about 420 wet tons/day of

RDF are required. Considering the RDF fraction in refuse and assuming the refuse facility to be open 5 days per week, about 750 wet tons of refuse would have to be received at the plant every day. Any commitment to use of refuse must recognize the relatively large quantities required to achieve autogenous combustion.

Apart from the cost, one other advantage of employing the C-G process in an energy recovery scheme is the elimination of any refuse handling needs at the wastewater treatment plant.

To combine the economic advantage of System 3 (C-G drying) with the higher revenue potential of System 6 (refuse addition) System 7 was developed. This system would comprise of dehydration of sludge before blending the same amount of refuse as in System 6 for thermal processing. The cost and energy analyses are shown in Table 5.8. The summary of the system costs and energy balances are given in Tables 5.9 and 5.10, respectively.

## MASS AND ENERGY BALANCE FOR SYSTEM 3

A mass and energy balance for the combined system of CG drying and thermal processing is shown in Figure 5.6

In developing the mass and energy balance it was assumed that any extracted heavy oil from the CG process is combined with dry product solids and pyrolyzed in the multiple hearth furnace. Low BTU fuel gas produced under partial oxidation conditions in the multiple hearth would be immediately afterburned and passed through a boiler for steam production. The mass flow and temperature of flue gases exiting the afterburner were calculated from a heat and mass balance considering the infeed caloric content and 140 percent stoichiometric air supply. Steam production was calculated assuming an exit gas temperature from the boiler of 400°F. Electrical production would be by condensing steam turbine with an extraction of process steam for the CG unit at 165 psia. Other power systems have been proposed including direct combustion of the fuel gas in a combined cycle system. The latter would be electrically more efficient than the system shown in Figure 5.6. However, it has not yet been demonstrated in full scale practice with the low BTU gases produced in a multiple hearth furnace. Therefore, the more proven, but perhaps less efficient system shown in Figure 5.6 was assumed for purposes of analysis.

About 38 percent of the steam produced is required for operation of the CG process with the remainder available for production of electricity. About 6260 KW of the total 7580 KW produced results from steam condensed at 3.2 psia. The remainder is derived from that portion of input steam extracted at 165 psia. If vacuum hydroextraction is used only about 22 percent of the process steam would be required at 165 psia with the remainder at 80 psia. If two extractions were made from the turbine, one each at 80 and 165 psia, electrical production could be increased to about 8019 KW, approximately a 6 percent increase. It should be noted that other

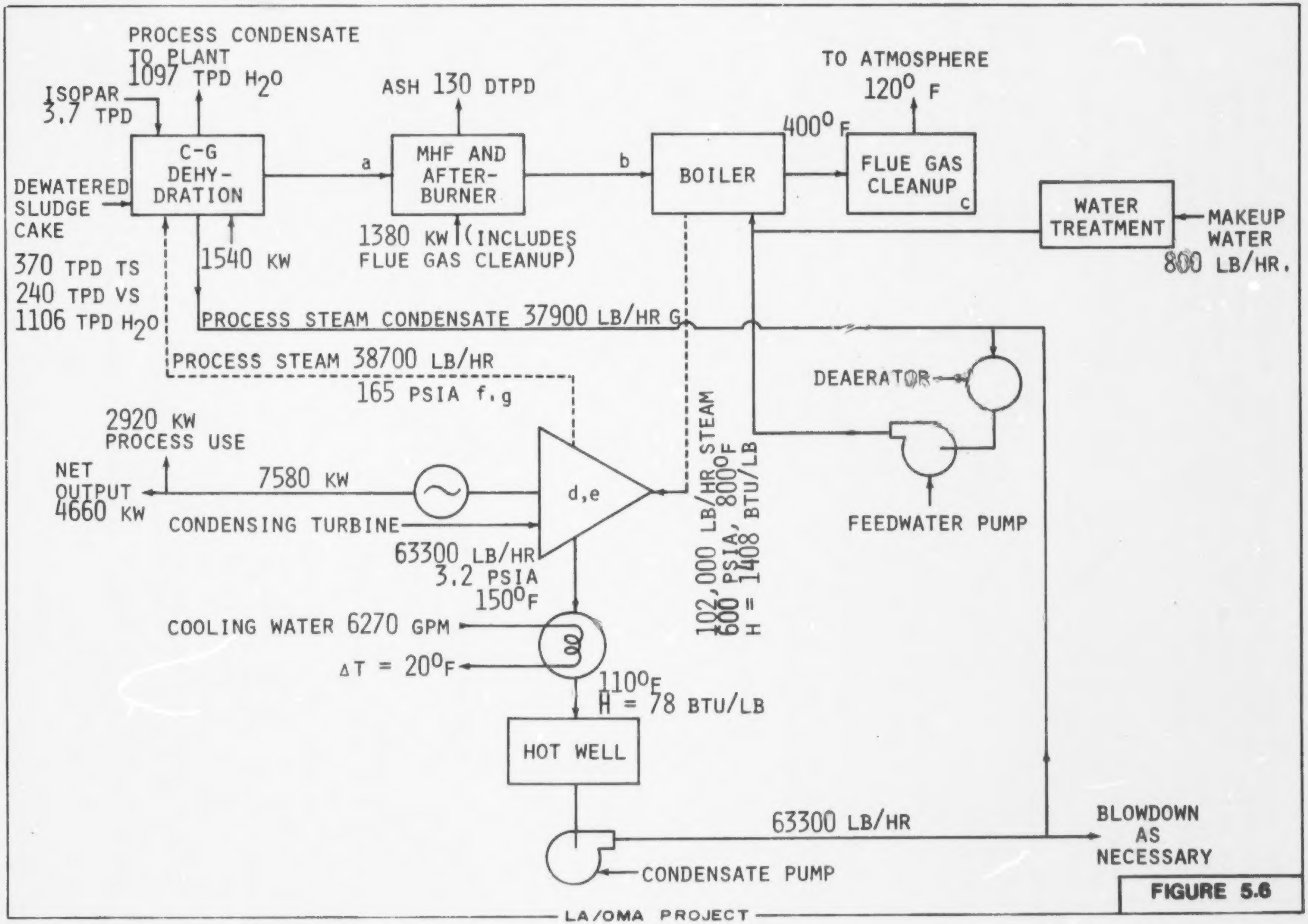

PROCESS CONDENSATE
TO PLANT
1097 TPD $H_2O$
ISOPAR
3.7 TPD
C-G DEHY-DRATION
DEWATERED SLUDGE CAKE
370 TPD TS
240 TPD VS
1106 TPD $H_2O$
1540 KW
a
MHF AND AFTER-BURNER
ASH 130 DTPD
1380 KW (INCLUDES FLUE GAS CLEANUP)
b
BOILER
400° F
FLUE GAS CLEANUP
c
TO ATMOSPHERE
120° F
WATER TREATMENT
MAKEUP WATER
800 LB/HR.
PROCESS STEAM CONDENSATE 37900 LB/HR G
PROCESS STEAM 38700 LB/HR
165 PSIA f,g
DEAERATOR
FEEDWATER PUMP
102,000 LB/HR STEAM
600 PSIA, 800°F
H = 1408 BTU/LB
2920 KW
PROCESS USE
NET OUTPUT
4660 KW
7580 KW
d,e
CONDENSING TURBINE
63300 LB/HR
3.2 PSIA
150°F
COOLING WATER 6270 GPM
ΔT = 20°F
110°F
H = 78 BTU/LB
HOT WELL
63300 LB/HR
CONDENSATE PUMP
BLOWDOWN AS NECESSARY
LA/OMA PROJECT

FIGURE 5.6

a/ Feed to MHF contains 217 dtpd of VS @ 9100 BTU/lb VS combined with 23 tpd of extracted heavy oil @ 18,000 BTU/lb and 18 tpd of residual water.

b/ Mass Flow and Temperature calculated from heat balance assuming 140 percent stoichiometric air in MHF and afterburner combined. Flue gas temperature is greater than 2800$^{\circ}$F in this example.

c/ High Energy venturi scrubber and multiple tray scrubber assumed.

d/ Steam turbine electrical production at 10.1 lb/hr per KW extracted @ 3.2 psia and 29.4 lb/hr per KW extracted @ 165 psia.

e/ Separate backpressure turbine could also be used.

f/ With vacuum applied to hydroextraction unit, about 22 percent of steam requirement would be at 165 psia with remainder at 80 psia. Two extractions from turbine at those pressures would be possible with some increase in electrical conversion efficiency.

g/ Assumes 5 percent consumptive loss of steam in vacuum hydroextraction and Isopar distillation system. If hydroextraction conducted at one atmosphere, consumptive loss could increase to 65 percent of applied steam.

**FIGURE 5.6 (CONT.)**
CARVER-GREENFIELD-
THERMAL PROCESSING SYSTEM
MASS AND ENERGY BALANCE

steam turbine configurations are possible for example, a condensing turbine with a separate backpressure turbine for steam used in the Carver-Greenfield. However, electrical production would not differ significantly. Therefore, the system shown in Figure 5.6 is felt to adequately represent the potential energy budget for a combined process of dehydration and thermal processing using a proven energy conversion system.

Net electricity produced from the system employing CG process is estimated at about 112,000 KWH/day (Table 5.4/Figure 5.4). By comparison, the system utilizing RDF (R/S=1.6) would generate about 158,000 KWH/day of surplus electricity. About 253,000 KWH/day of electricity is estimated from a combined cycle power plant using digester gas produced on the same quantity of raw sludge. Thus dehydration and thermal processing can produce net electricity in comparable amounts to the system using dewatered cake and refuse at R/S=1.6. However, both systems would produce less net energy than that available from the digester gas. This is due to the fact that digester gas can be used in more efficient energy conversion systems. Efforts are currently being made to use low BTU pyrolytic gas in combined cycle power systems. If these efforts are successful, electrical energy production from both Systems 3 and 6 could be significantly improved.

# APPENDICES

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# APPENDIX I

## ISOPAR L

## DESCRIPTION, SPECIFICATION AND TEST PROCEDURE

# ISOPAR® L *

## Description

Isopar L is a carefully fractionated isoparaffinic hydrocarbon of exceptional purity with a mid-boiling point approximately 381°F. Produced by modern synthesis from selected petroleum fractions, Isopar L is largely composed of saturated paraffins of highly branched molecular structures. Stringent manufacturing specifications limit trace impurities—sulfur, acids, carbonyls, chlorides, etc.—to a few parts per million. Isopar L is essentially odorless, and has excellent odor stability.

## Applications

In addition to its primary application in odorless paint formulations, the inherent characteristics of Isopar L lend themselves to numerous solvent and process applications, and to the formulation of a variety of proprietary compounds.

## Specification Properties

| | Minimum | Maximum | Test Method |
|---|---|---|---|
| Gravity, °API, 60/60°F | 52.5 | 54.5 | ASTM D 287 |
| Color, Saybolt | 30 | — | ASTM D 156 |
| Aniline point, °F | 183 | — | ASTM D 611 |
| Sulfur, ppm | — | 10 | ASTM D 1266 (Nephelometric mod.) |
| Distillation, °F | | | ASTM D 86 |
| IBP | 370 | 380 | |
| 50% | 378 | 385 | |
| Dry point | 400 | 410 | |
| FBP | — | 415 | |

## Typical Properties

| | | Test Method |
|---|---|---|
| **Solvency** | | |
| Aniline point, °F | 187 | ASTM D 611 |
| Solubility parameter | 7.3 | Burrell(1) |
| Kauri-butanol value | 27 | |
| **Miscibility** | | |
| With other petroleum solvents | all proportions | |
| With chlorinated hydrocarbons | all proportions | |
| With alcohols | (2) | |
| **Volatility** | | |
| Distillation | | |
| IBP | 373 | ASTM D 86 |
| 5% | 375 | |
| 10% | 376 | |
| 50% | 381 | |
| 90% | 393 | |
| 95% | 400 | |

| | | Test Method |
|---|---|---|
| **Volatility cont.** | | |
| Dry point | 406 | |
| FBP | 413 | |
| Flash point, °F | 144 | ASTM D 56 |
| Vapor pressure, psia | | Calculated |
| at 150°F | 0.10 | |
| at 200°F | 0.48 | |
| at 250°F | 1.6 | |
| Non volatile content, mg/100ml | 0.7 | ASTM D 268 |
| Evaporation rate, seconds | | Federation of Societies for Paint Technology Method II |
| 10% | 660 | |
| 30% | 2400 | |
| 50% | 4680 | |
| 70% | 7560 | |
| 90% | 11280 | |
| 95% | 12420 | |
| 100% | 13680 | |

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(1) Federation of Societies for Paint Technology, **Official Digest**, October, 1955, p. 726

(2) Miscible in all proportions with alcohols above methanol, with ketones, and with pyridine; immiscible with polyhdric alcohols; limited miscibility with methanol and pyrole

| **Composition** | | **Test Method** |
|---|---|---|
| Average molecular weight | 165 | Calculated |
| Hydrocarbon type, vol % | | Mass Spectrometer |
| Normal paraffins | <0.05 | |
| Isoparaffins | 92.7 | |
| Total paraffins | 92.7 | |
| Naphthenes: | | |
| 1-ring | 7.0 | |
| 2 ring | 0.0 | |
| Total naphthenes | 7.0 | |
| Mono aromatics | 0.2 | |
| Olefins | 0.06 | Calculated from Br. No.—ASTM D 1491 |
| Unsulfonated residue, vol % | 99.7 | ASTM D 483 |
| Trace compounds | | |
| Sulfur | | |
| Sulfides (Doctor test) | pass | ASTM D 484 |
| Total sulfur, ppm | 5 | ASTM D 1266 (Nephelometric mod.) |
| Peroxides, mg/liter active oxygen | <1 | |
| Inorganic acids | nil | |
| Organic acids | nil | |
| Carbonyl number, mg KOH/g | <0.01 | |
| Chlorides, ppm | <1 | |
| Nitrogen, ppm | <1 | |
| Bromine index | 120 | ASTM D 1491 |
| Odor, bulk | ~none | |
| Odor, residual | none | |
| Odor stability | excellent | |
| **Surface Properties** | | |
| Surface tension, dynes/cm at 25°C | 22.9 | |
| Interfacial tension, dynes/cm at 25°C | 49.2 | |
| Spreading coefficient | 1.0(3) | |
| Demulsibility, ml/hr | 1620 | |

| **Surface Properties cont.** | | **Test Method** |
|---|---|---|
| Work of cohesion, ergs/cm$^2$ | 45.8(4) | |
| Work of adhesion, ergs/cm$^2$ | 44.8(5) | |
| **Water Solubility at 25°C** | | |
| Water in hydrocarbon, ppm | 103 | Karl Fisher |
| Hydrocarbon in water, ppm | <0.001 | Calculated |
| **Health Hazards** | | |
| Inhalation, MAC, ppm | 500 | |
| Oral LD$_{50}$, ml/kg | >10 | |
| Skin LD$_{50}$, ml/kg | >3 | |
| **General Properties** | | |
| Gravity, °API, 60/60°F | 52.9 | ASTM D 287 |
| Gravity, specific, 60/60°F | 0.7674 | |
| Pounds per gallon, 60/60°F | 6.388 | |
| Density, at 5°C | 0.772 | |
| at 25°C | 0.758 | |
| at 40°C | 0.746 | |
| at 60°C | 0.733 | |
| Refractive index, 20°C | 1.42735 | |
| Viscosity, | | |
| cp at 25°C | 1.600 | |
| cp at 100°F | 1.287* | |
| cs at 100°F | 1.667 | |
| Color, Saybolt | 30 | |
| Color stability, 16 hr at 212°F, Saybolt | 30 | |
| Specific heat, liquid, Btu/lb/°F | | |
| at 60°F | 0.466 | |
| at 150°F | 0.529 | |
| at 200°F | 0.560 | |
| Heat of vaporization, Btu/lb | | |
| at 212°F | 122 | |
| at BP (atmos. press.) | 108 | |
| Freezing point | <−76°F | |

(3) Spreading coefficient = surface tension of water − (surface tension of hydrocarbon + interfacial tension hydrocarbon/water)
(4) Work of cohesion = 2 (surface tension of hydrocarbon)
(5) Work of adhesion = surface tension of hydrocarbon + (surface tension of water — interfacial tension hydrocarbon/water)

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* As per Technical Brochure (1965) from

**HUMBLE OIL & REFINING COMPANY**
Houston, Texas

## LABORATORY ANALYSIS a/

Object - Method for the determination of Isopar H

Scope - Applicable to samples taken from solids dried in Insopar H and containing more than 1.0% Isopar H.

Method - This metod used the apparatus of A.O.C.S. official method Ca. 2a-45, i.e., which is used for the determination of moisture making use of the Dean-Stark trap to collect the Isopar through steam distillation.

1. Weigh 50.0 grams of solids in a 500 cc extraction flask.

2. Add 100 cc water.

3. Add 75 cc of refine and deodorized vegetable oil such as Wesson oil and twirl to mix.

4. Place the flask in a hot plate and condense and collect the Isopar H and water condensate in the calibrated trap that has a stop cock valve at the base of the trap. Periodically discharge the collected water leaving behind the Isopar H which is the upper phase.

5. When 100 cc of water has been collected, cool the trap containing the Isopar H and read the volume in milliters.

6. Calculations

    a. ML of Isopar x 0.75 (density) = gm Isopar H

    b. $\frac{\text{gm Isopar H}}{\text{gm sample}}$ = % Isopar in sample

a/ Supplied by Dehydrotech Corporation

# APPENDIX II

# MASS AND ENERGY BALANCE
# FOR A
# FOUR-EFFECT CARVER-GREENFIELD PROCESS

Prepared By:

Thomas Mize, Project Engineer
Treatment Plant Design
County Sanitation Districts of Los Angeles County

# HEAT AND MASS BALANCE FOR CARVER-GREENFIELD PROCESS

Capacity 20 Dry Tons/Day Feed Sludge @ 20% T.S.

It will be assumed that condensate does give up heat other than heat of vaporization for water and oil. This will mean that temperature of condensate out of Hex. will be the same temperature as vapor out of that stage. This will generally be a temperature drop of 25 degrees.

## Temperature of Recycle Oil

There is 14,200 lb. of oil and 1667 lb. of solids at 255°F returned as recycled oil. This is mixed with 5000 lb. of oil at 210°F that is returned from third stage heat exchanger.

## Mix Temperature

$$(14{,}200)\ (.55)\ (255-T)+(1667)\ (.3)\ (255-T) = (5000)(.55)(T-210)$$

$$(M_{oil})\ (C_{oil})\ (\Delta T)\ +(M_{solid})(C_{solid})(\Delta T) = (M_{oil})(C_{oil})(\Delta T)$$

$$1{,}991{,}550 - 7810T + 127{,}525 - 500T = 2750T - 577{,}500$$

$$11060\ T = 2{,}696{,}575$$

$$T = 243.8^{\circ}F$$

$$\text{USE } T = 240^{\circ}F$$

## Temperature of Mixture

$$(20{,}000)(.55)(130-T)+(1667)(.3)(130-T)=(6667)(1)(T-70)= (1667)(.3)(T-70)$$

$$1{,}430{,}000 - 11{,}000T + 65{,}000 - 500T = 6667T-466{,}690+500T-35{,}000$$

$$18{,}667T = 1{,}996{,}690$$

$$T = 107^{\circ}F$$

For heat balance calculations, use 32°F as reference

## Heat Content of Feed

Water and solids at 70°F

For Water

$Q = MC\Delta T = (6667)(1)(70-32) = 253{,}346$ BTU/HR.

For Solid

$Q = MC\Delta T = (1667)(.3)(70-32) = \underline{19{,}000}$ BTU/HR.

TOTAL: 272,346 BTU/HR.

## Heat Content of Recycled Oil

Oil and Solids at 130°F

**Oil**

$Q = MC\Delta T = (20{,}000)(.55)(130-32) = 1{,}078{,}000$

**Solid**

$Q = MC\Delta T = (1667)(.3)(130-32) = \underline{49{,}000}$

1,127,000 BTU/HR

## Heat Content of Fluidized Feed and Oil

272,346 + 1,127,000 = 1,399,346 BTU/HR

As indicated in the Carver-Greenfield calculations, there is 1950 lb. of water vaporized in each of the first three stages and 817 lb. of water vaporized in the fourth stage. For each lb. of water vaporized, there will be 0.4 lb. of Isopar Oil vaporized.

## First Stage

This stage will have a vacuum of 26.475 In, Hg which will reduce boiling point of $H_2O$ to 120°F.

- Heat requirement to vaporize $H_2O$ + oil

Oil $Q = (780)(120) = 93{,}600$
$H_2O$ $Q = (1950)(1000) = \underline{1{,}950{,}000}$

TOTAL: 2,043,600 BTU/HR

- Heat in vapor from temperature increase (32 → 120)

Oil $Q = (780)(.55)(120-32) = 37{,}750$
$H_2O$ $Q = (1950)(1)(120-32) = \underline{171{,}600}$

TOTAL: 209,350 BTU/HR

TOTAL HEAT IN VAPOR FROM 1st STAGE

2,043,600 + 209,350 = 2,252,950 BTU/HR

- Heat Content of Fluid Leaving 1st Stage

Oil $Q = (19{,}220(.55)(120-32) = 930{,}248$
Solid $Q = (3{,}333)(.3)(120-32) = 87{,}991$
$H_2O$ $Q = (4{,}717)(1)(120-32) = \underline{415{,}096}$

TOTAL: 1,433,335 BTU/HR

- Heat in vapor to first stage (excluding heat of vapor)

Oil Q = (780)(.55)(145-32) = 48,477
$H_2O$ Q = (1950)(1)(145-32) = 220,350
268,827 BTU/HR

Heat of vaporization of oil
+ $H_2O$ 2,043,600
2,312,427 BTU/HR

- Heat in condensate + oil out of 1st HEX.

From Above: $Q_{oil}$ = 37,750 BTU/HR
$Q_{H_2O}$ = 171,600 BTU/HR
TOTAL: 209,350 BTU/HR

SUMMARY OF HEAT IN STAGE 1

| Heat In | | Heat Out | |
|---|---|---|---|
| Slurry Feed | - 1,399,346 | Slurry Feed | - 1,433,335 |
| Vapor from St. 2 | - 2,312,427 | Vapor Provided | - 2,252,950 |
| Flashing | - 50,000 | Oil in Condensate | - 37,750 |
| | | Condensate | - 171,600 |
| | 3,761,773 | | 3,985,635 |

- 133,862 more heat out

This means that we are not vaporizing 1950 lb. of water in this stage (1st)

∴ 1816 lb/ of water is vaporized

Second Stage

- Heat rejection from recycled oil at HEX #3

$Q_{oil}$ = (19,200)(.55)(155-30) = 264,000 BTU/HR

$Q_{solids}$ = (1667)(.55)(155-130) = 12,500 BTU/HR

TOTAL: 276,500 BTU/HR

- Total heat content of slurry to 2nd stage

1,433,335 + 276,500 = 1,709,835 BTU/HR

- Heat content of slurry leaving 2nd stage

$Q_{oil}$ = (19,200)(.55)(145-32) = 1,193,280

$Q_{solids}$ = (3,333)(0.3)(145-32) = 113,000

$Q_{H_2O}$ = (2,761)(1)(145-32) = 311,993

TOTAL: 1,618,273 BTU/HR

- Heat content of vapor to 2nd stage

(from page 3)Heat of vaporization (1950#$H_2O$,780#oil)=2,043,600 BTU/HR

Heat in vapor from temperature increase (170-32)
oil Q = (780)(.55)(170-32) = 59,202
$H_2O$ Q = (1950)(1)(170-32) = 269,100 BTU/HR
Heat of vaporization 2,043,600

TOTAL: 2,371,902 BTU/HR

- Heat content of condensate + oil from 2nd stage

oil Q = (780)(.55)(145-32) = 48,477 BTU/HR
$H_2O$ Q = (1950)(1)(145-32) = 220,350 BTU/HR

SUMMARY OF HEAT IN STAGE 2

| Heat In | | Heat Out | |
|---|---|---|---|
| Slurry Feed | - 1,709,835BTU/HR | Slurry Feed | - 1,618,273BTU/HR |
| Vapor from St. 3 | - 2,371,900 | Vapor Produced | - 2,313,427 |
| Flashing | - 50,000 | Oil in Condensate | - 48,477 |
| Oil | - 37,750 | Condensate | - 220,350 |
| | 4,169,485BTU/HR | | 4,199,527BTU/HR |

- 30,042 BTU/HR more heat out

∴ 1920 lb. of water is vaporized

Third Stage

- Heat content in slurry feed leaving 3rd stage

oil - Q = (19,220)(.55)(185-32) = 1,617,363 BTU/HR
solids - Q = (3,333)(.3)(185-32) = 152,984
$H_2O$ - Q = (817)(1)(185-32) = 125,000

TOTAL: 1,895,347 BTU/HR

- **Heat content in slurry feed to 3rd stage**

Heat rejection at HEX #2 from recycled oil
oil Q = (19,220)(.55)(195-155) = 422,840
solids Q = (1667)(.3)(195-155) = 20,000

Slurry from 2nd stage (from page 4) 1,618,273 BTU/HR

TOTAL: 2,061,113

- **Heat content of vapors to 3rd stage HEX.**

1. It is assumed that 5000 lb. of oil vapor at 325°F will pass through 4th stage vapor chamber without heat loss.

   oil Q = (5000)(.55)(325-32) = 805,750 BTU/HR (to heat to 325°F)
   Q = (5000)(120) = 600,000
   Oil from still 1,405,750 BTU/HR

2. Isopar Oil vaporized in 4th chamber has temperature of 210°F

   oil Q = (817)(.55)(210-32) = 79,984 To Heat
   Q = (817)(120) = 98,000 To Vaporize
   177,984 BTU/HR

3. It will be assumed that the 500 lb. of steam from the hydroextractor will remain at 325°F until it reaches the 3rd stage HEX.

   $H_2O$ Q = (500)(1)(325-32) = 146,500 BTU/HR
   Q = (500)(1000) = 500,000
   646,500 BTU/HR

4. Heat in the 817 lb. of steam from 4th stage

   $H_2O$ Q = (817)(1)(210-32) = 145,426 BTU/HR Heating Vapor
   Q = (817)(1000) = 817,000
   962,426 BTU/HR

Summing the above numbers from items 1 through 4

1. 1,405,750 BTU/HR
2. 177,984
3. 646,500
4. 962,426

→ 3,192,660 BTU/HR in vapor to 3rd stage HEX.

- **Heat content in condensate and oil out of 3rd stage HEX**

oil: Q = (5000)(.55)(210-32) = 489,500
$H_2O$ Q = (1318)(1)(210-32) = 234,600
TOTAL: 724,100 BTU/HR
oil: Q = (817)(.55)(210-32) = 79,984

## SUMMARY OF HEAT IN STAGE 3

| Heat In | | Heat Out | |
|---|---|---|---|
| Slurry Feed | 2,061,113BTU/HR | Slurry | 1,895,347BTU/HR |
| Vapor from St. 4 | 3,192,660 | Vapor Produced | 2,371,900 |
| Oil | 48,477 | Oil | 79,984 |
| | 5,302,250BTU/HR | Condensate & Oil | 724,100 |
| | | | 5,071,331BTU/HR |

+ 231,000 BTU/HR

The additionally heat in means we can vaporize more than 1950 lb. of water in the 3rd stage HEX. we have a carry over of 133 lb. of $H_2O$ from Stage 1 and a carry over of 30 lb. of water from Stage 2.

∴ We have excess water 133 + 30 = 163 lb. and excess heat to vaporize 231 lb. The net result is that there is approximately 70,000 BTU/HR excess heat. This heat will go into the slurry feed to increase the temperature or will be contained in the condensate flow. Up through the first three stages, there is fairly good agreement.

### Fourth Stage

- Heat content in slurry leaving 4th stage

oil: Q = (19,220)(.55)(255-32) = 2,357,333
solids Q = (3,333)(.3)(255-32) = 222,977
No Water
2,580,310 BTU/HR.

- Heat content in slurry feed to 4th stage

Heat Rejection at HEX #a from recycled oil

oil: Q = (19,220)(.55)(240-195) = 475,695 BTU/HR
solids: Q = (1,667)(.3)(240-195) = 22,500
BTU/HR from 3rd stage = 1,895,347
2,393,542 BTU/HR

- Determination of steam requirement at stage 4

| Heat In | | Heat Out | |
|---|---|---|---|
| Slurry feed | 2,393,542BTU/HR | Slurry | 2,580,310BTU/HR |
| Vapor from still | 2,052,250 | Vapor | 3,192,660 |
| Oil | 79,984 | | |
| Steam | X | | |
| | 4,525,776 + X | | 5,772,970BTU/HR |

4,525,776 + X = 5,772,970

X = 1,247,194 BTU/HR

Steam at $270^{o}F$ and 27 psig has 945 BTU/lb available for heating ($h_{fg}$ = 931 BTU/lb plus 14 BTU/lb temperature decrease → 945 BTU/lb.)

$$\text{Quantity of steam requirement} = \frac{1{,}247{,}194 \text{ BTU/HR}}{945 \text{ BTU/LB}} = 1320 \text{ lb/hr steam}$$

→ Input to 4th stage:  1320 lb/hr steam

Hydroextractor

The feed from the 4th stage consist of 19,220 lb. oil and 3,330 lb. solids. 1667 lb. of solids are returned to fluidizing tank to improve mixing. This bleed off of solids takes 9600 lb. of oil. The remaining 9733 lb. (oil + heavy oil) and 1534 lb. of solids goes to the centrifuge which produces two streams. One stream consists of 8733 lb. of light + heavy oil, and a second stream consists of solids and oil (1000 lb/hr of oil and 1534 lb/hr of solids). This second stream goes to the hydroextractor. It will be assumed that the solids contain 4 percent moisture. There is a direct injection of 500 lb. of steam into the hydroextractor to reduce the boiling point due to effect of partial pressure from the steam.

Heat In

oil: Q = (1000)(.55)(255-32) = 122,650
solids: Q = (2534)(.3)(255-32) = 102,624
$H_2O$: Q = (500 lb)(1000 BTU/LB)= 500,000
Q = (X lb)(900 BTU/LB) = 900X
(726,174 + 900X) BTU/HR

Heat Out

oil: Q = (1000)(.55)(325-32) = 161,150
Q = (1000)(120) = 120,000
solids: Q = (1534)(.3)(325-32) = 134,838
steam: Q = (500lb/hr)(1000BTU/LB)=500,000
$H_2O$ Q = (1534)(0.03)(1000) = 46,000

961,988 BTU/HR

726,174 + 900X = 961,988

900X = 226,714 BTU/HR

X = 262 lb. steam need

Total steam requirement:

500 lb. for direct injection
262 lb. for heating

Total: 762 lb.

Separation Still

Heat Out

Oil + steam vapor Q = 2,052,250 BTU/HR

Heavy oil Q = (133)(.55)(325-32) = 21,432 BTU/HR

Condensate Q = (X)(1)(325-32) = 293X
Total: 293X + 2,073,682

Heat In

- Vapor from Hydroextractor:

oil: Q = (1000)(.55)(325-32) = 161,150 BTU/HR
Q = (1000)(120) = 120,000
$H_2O$: Q = (500)(1)(325-32) = 146,500
Q = (500)(1000) = 500,000

- Recycled Oil:

(Isopar) Q (4000)(.55)(255-32) = 490,600
(heavy) Q (133)(.55)(255-32) = 16,312

- Steam Q = 900X + 293X 1,193X

TOTAL: 1193X + 1,434,562

$$293X + 2,073,682 = 1193X + 1,434,562$$

$$900X = 639,120$$

$$X = 710 \text{ lb. steam}$$

Summary of steam requirement for 20 dry tons per day
20% solids

To 4th Stage - 1320 lb. (270°F 27 psig)
To Hydroextractor - 762 lb. (500 lb. direct; 252 lb. indirect)
(335°F, 94 psig)
To Still - 710 lb. (335°F, 94 psig)

TOTAL: 2792 lb.

Accounting for 5% heat loss: TOTAL = 2930 lb/hr

- Total Steam Required = 2930 lb./hr (with 5% heat loss)
- Solids Content = 20 tons/day (20% solids)
  = 1667 lb/hr
- Water vaporized = 6667 lb/hr
- Amount of steam required per pound of water vaporized

$$\frac{2930 \text{ lb/hr steam}}{6667 \text{ lb/hr } H_2O \text{ var.}} = 0.439 \frac{\text{lb steam}}{\text{lb } H_2O} \text{ (with 5\% heat loss)}$$

Based on data from the Coors installation, the steam requirement is 0.347 lb steam/lb $H_2O$, but Coors does not utilize a hydroextractor or separation still. They get the full benefit of the four effects.

∴ 0.439 lb steam/lb $H_2O$ vaporized is a reasonable number

- Amount of steam required per pound of solid.

$$\frac{2928 \text{ lb/hr steam}}{1667 \text{ lb/hr solids}} = 1.75 \frac{\text{lb steam}}{\text{lb solid}} \text{ (for 20\% solids)}$$

- Allocation of steam to individual components

1. Fourth Stage Evaporator - 1320 lb → 48%
2. Hydroextractor - 752 lb → 27%
3. Separation Still - 710 lb → 25%

2782 lb

- Output of centrifuge

1000 lb of oil and 1534 lb of solids

∴ 60% solids

# APPENDIX III

## ASSUMED DESIGN CRITERIA AND PROJECTED ENERGY BALANCE

# DESIGN CRITERIA FOR CARVER-GREENFIELD DEHYDRATION

## INTRODUCTION TO PROCESS

Carver-Greenfield dehydration is a multiple effect evaporation scheme using Isopar-L (an oil) as the transport liquifying agent for the sludge. The equipment included in the process consists of 4-effect evaporators, pumps, steam generation facility, hydroextractors, and centrifuges.

## DESIGN CRITERIA AND PROCESS FACTORS

The design criteria that do not vary with process feed are listed in Table A. Their values are used irrespective of sludge feed. The recycle soluble COD and volatile solids reduction to recover heavy oil does vary with feed and is listed separately in Table B. Process feeds investigated were:

- Belt Press Dewatered Thickened Waste Activiated Sludge
- Belt Press Dewatered Digested Thickened Waste Activated Sludge
- Belt Press Dewatered Digested Combined Primary Sludge and Thickened Waste Activated Sludge

## PROCESS FACTORS DERIVATION

### Capital Cost

All capital cost information was supplied by Carver-Greenfield. Size scaling was done by a curve fit using the evaporative burden as the independant variable. Process factors which were used to compute capital costs are listed in Table C.

### Operation and Maintenance Cost

The O&M cost information was also supplied by Carver-Greenfield. Table D lists factors used to compute O&M costs.

### Energy Factors

Electrical power and fuel consumption estimates were obtained by evaluating data obtained from Carver-Greenfield. Fuel consumption was determined by calculating the total fuel required and then subtracting the fuel produced in the process. Table E lists factors used to compute electrical energy and fuel consumption.

### Air Emission Factors

Estimate of air emission are based on the burning of low sulfur fuel oil and/or heavy oil from process. The emissions are controlled by tray scrubbers. Factors used are listed in Table F.

## TABLE A

## CARVER-GREENFIELD DEHYDRATION DESIGN CRITERIA

| | |
|---|---|
| Number of Effects | 4 |
| Evaporation Rate | 2.3 lbs water/lb steam |
| Steam Characteristics | 50 psig saturated |
| Boiler Efficiency | 75% |
| Redundancy | 3-4 effect modules, each module able to handle one-half the evaporative burden. |
| Fuel Value of Extracted Heavy Oil | 18,000 Btu/lb |
| Fluidizing Oil | Isopar-L (an Exxon product) |
| Fluidizing Oil Make-Up | 1% by weight of dry solids fed |
| Weight of Isopar-L | 7.5 lbs/gal |
| Outfeed | 95% solids |

## TABLE B

## CARVER-GREENFIELD DEHYDRATION PARAMETERS THAT VARY WITH SLUDGE FEED

| | VSS Reduction % * | Recycle COD mg/l | Recycle $O_2$ lbs/lb COD |
|---|---|---|---|
| Belt Press Dewatered Thickened Waste Activated Sludge (WAS) | 5 | 2,000 | 0.45 |
| Belt Press Dewatered Digested WAS | 5 | 2,000 | 0.45 |
| Belt Press Dewatered Digested Combined Primary Sludge and WAS | 10 | 2,500 | 0.48 |

* Extracted as heavy oil

## TABLE C

## CARVER-GREENFIELD DEHYDRATION CAPITAL COST FACTORS

| | |
|---|---|
| Evaporator Cost ($) | $3 \times (2.8(10^6) + (4{,}600 \times (\frac{\text{Evap.Burden}}{2}\ \text{t/d})))$ |
| Boiler Cost ($) | $3 \times (((\frac{\text{Steam req'd t/d}}{2}) \times \frac{2{,}000\ \text{lb/t}}{24\ \text{hrs/day}})$<br>$\times\ \$8/\text{lb/hr}$ |
| Building Size* | $22{,}500\ \text{ft}^2 (\text{Evap. Burden } (\text{t/d})/1{,}440)^{1/2}$ |
| Building Cost | $\$74 \text{ per ft}^2$ |
| Gross Land Area | 1.9 x building size |
| Contingency | 40% |

ENR = 3031

* Entire system housed

## TABLE D

## CARVER-GREENFIELD DEHYDRATION OPERATION AND MAINTENANCE COST FACTORS

| | |
|---|---|
| Labor | 8 positions regardless of size |
| Maintenance | 3 percent of annual process capital cost |
| Fuel Cost | \$0.40 per gallon |
| Electrical Cost | \$0.03 per kWh |
| Recycle $O_2$ Cost | \$22.00/ton $O_2$ required |
| Isopar-L Cost | \$0.70 per gallon |
| Heavy Oil Credit | \$0.40 per gallon |

CPI = 183

## TABLE E

## CARVER-GREENFIELD DEHYDRATION ENERGY USE AND PRODUCTION FACTORS

| | |
|---|---|
| Electric Power | 33 kWh/ton water evaporated |
| Electric Power for Recycle $O_2$ | 13 kWh/ton $O_2$ |
| Fuel Usage for Boiler | |
| Steam Enthalpy* | 3.07 $(10^6)$ Btu/ton of steam |
| Heat Value of Fuel | 37.0 $(10^6)$ Btu/ton of fuel |
| Fuel Required | $\frac{3.07\ (10^6)}{37\ (10^6)} = 0.083\ \frac{\text{t/d fuel}}{\text{t/d steam}}$ |

* Includes boiler efficiency at 75%

## CARVER-GREENFIELD DEHYDRATION AIR EMISSION FACTORS

| | |
|---|---|
| Particulate | 0.29#/dry ton stack gas |
| $NO_x$ | 0.30#/dry ton stack gas |
| $SO_2$ | 4.0#/ton low sulfur fuel oil burned<br>20#/ton heavy oil burned |

# DESIGN CRITERIA FOR THERMAL PROCESSING IN A MULTIPLE HEARTH FURNACE

## INTRODUCTION TO PROCESS

Direct Fired Thermal Processing, for the purposes of this report is defined as partial combustion of dewatered sludge and (in some cases) refuse derived fuel (RDF) in a multiple hearth furnace using less than stoichiometric amounts of air and immediate combustion of resultant offgasses using stoichiometric amounts of air in an afterburner. Included in the process are heat recovery, steam production, and electrical power generation. Also included are emission control to reduce air pollution using high energy loss venturi scrubbers followed in series by multiple tray wet type impingement scrubbers. In all cases, afterburner outlet temperature is raised to a minimum 1400°F temperature through the burning (if required) of either RDF or supplemental fuel oil in the system.

## DESIGN CRITERIA AND PROCESS FACTORS

Pertinent design criteria used in the engineering and economic analyses of sludge management systems involving the use of Direct Fired Thermal Processing are listed in Table F. The design criteria listed therein, except as noted in Table G, are uniformly used regardless of process feed to Direct Fired Thermal Processing. Process feeds investigated were:

- Belt Press Dewatered, Gravity Thickened, Thermally Conditioned, Thickened WAS plus Belt Press Dewatered Digested Primary Sludge
- Solar Dried, Digested Combined Primary Sludge and Thickened WAS

- Dehydrated, Belt Press Dewatered, Digested, Thickened WAS
- Dehydrated, Belt Press Dewatered, Digested, Combined PS and Thickened WAS
- Belt Press Dewatered, Gravity Thickened, Thermally Conditioned, Combined PS and Thickened WAS
- RDF (2 ratios) plus Belt Press Dewatered, Digested, Combined PS and Thickened WAS
- Dehydrated, Belt Press Dewatered, Thickened WAS plus Belt Press Dewatered Digested PS

Each of these process feeds exhibits different characteristics as regards feed moisture content, feed volatile and nonvolatile content, and Btu value. These variables were computed and subsequently used in the analysis of Direct Fired Thermal Processing. These variables directly offset certain performance criteria, including:

- Afterburner temperature.
- Boiler efficiency and quantity of steam produced.
- Amount of power produced.
- Amount and character of stack gasses and pollutants emitted.

In turn, such performance criteria, along with the design criteria assumed and the amount of sludge to be processed, significantly offset factors relating to such parameters as:

- Capital and O&M Cost
- Energy Use and Production
- Air Emissions
- Land Requirements

## PROCESS FACTORS DERIVATION

### Capital Cost

Capital cost estimates for Direct Fired Thermal Processing are largely based on cost information supplied by Envirotech Corporation. Capital cost is a function of the number of furnaces required, size of heat recovery and power generation equipment, building requirements, land costs, and RDF storage and feed facilities when RDF fed. Process factors which were used in this analysis to compute capital costs are listed in Table H.

### Operation and Maintenance Cost

Labor requirements are based on information supplied by various sources(1) and are a function of the number of furnace units required. Table I lists factors used to compute O&M costs used in this analysis.

(1) Envirotech Corporation, Nichols Engineering Corporation, USEPA, CH2M HILL

## Energy Factors

Electrical power consumption and production estimates are based on information supplied by various sources (2) and are a function of the wet infeed load, Btu value of infeed, and equipment design. Table J lists factors used to compute electrical power consumption and production in this analysis.

## Air Emission Factors

Estimates of air emissions are based on the use of high energy loss venturi scrubbing followed in series by multiple wet tray type impingement scrubbers. Factors used to estimate the weight of various pollutants emitted are listed in Table K.

---

(2) Envirotech Corp., General Electric Co., Inc., CH2M HILL

## TABLE F

## DIRECT FIRED THERMAL PROCESSING DESIGN CRITERIA

| | |
|---|---|
| Fuel Value of Sludge | See following table |
| Fuel Value of RDF | 6,000 Btu/lb RDF (Wet Basis) |
| Size of Furnace | 25'9" (OD) x 12 hearth |
| Maximum Load per Furnace w/o redundancy | 650 tons per day wet basis total infeed |
| Redundancy | Rate units at 80% of maximum capacity and use minimum two furnaces. Round up number of furnaces to a whole number. |
| Characteristics of Steam Produced | 600 psig, 800°F, 1408 Btu/lb |
| Ash Produced | 0.15 lb ash/lb RDF (dry basis) plus 100% of nonvolatile content of sludge feed. |
| Minimum Afterburner Temperature | 1400°F |
| Air for Pyrolysis | 40% of Theoretical - 280 lbs air/MMBtu |
| Air for Afterburner | 100% of Theoretical - 700 lbs air/MMBtu |
| Emission Controls | High energy loss venturi plus multiple tray wet scrubber. |

## TABLE G

## DIRECT FIRED THERMAL PROCESSING PARAMETERS THAT VARY WITH SLUDGE FEED

| Sludge Infeed | Heat Value of Sludge Btu/lb VSS |
|---|---|
| Belt Press Dewatered, Gravity Thickened, Thermally Conditioned, Thickened WAS plus Belt Press Dewatered Digested Primary Sludge | 10,000 |
| Solar Dried, Digested Combined Primary Sludge and Thickened WAS | 10,000 |
| Dehydrated, Belt Press Dewatered, Digested, Thickened WAS | 9,100 |
| Dehydrated, Belt Press Dewatered, Digested, Combined PS and Thickened WAS | 9,100 |
| Belt Press Dewatered, Gravity Thickened, Thermally Conditioned, Combined PS and Thickened WAS | 10,000 |
| RDF (2 ratios) plus Belt Press Dewatered, Digested, Combined PS and Thickened WAS | 10,000 |
| Dehydrated, Belt Press Dewatered, Thickened WAS plus Belt Press Dewatered Digested PS | 9,600 |

## TABLE H

## DIRECT FIRED THERMAL PROCESSING CAPITAL COST FACTORS

| | |
|---|---|
| Number of Furnaces | Total wet feed (t/d)/(wet load per furnace (t/d) x redundancy factor), rounded upwards to a whole number. |
| Steam Produced | (Weight of stack gasses) x (temperature differential) x (specific heat of gasses)/(enthalpy change boiler feedwater to steam). |
| Temperature of Gasses (heat recovery) | 1,400°F minimum input, 400°F exit |
| Steam Turbines Water Rate | 10.1 lbs steam per kW (23% wire to water efficiency) |
| Cost per Furnace Unit | \$11,000,000 |
| Cost of RDF Storage and Feed | $(500 \times \text{wet RDF feed, t/d})^{0.5} \times$ \$1,000 |
| Cost of Elect. Generation | \$200 per kW |
| Building Size (for furnaces) | 8,125 $ft^2$ per furnace unit |
| Building Size (for power generation) | 0.35 $ft^2$ per kW |
| Building Cost | \$63.20 per $ft^2$ (5 level building) |
| Gross Land Area | 16,900 $ft^2$ per furnace + 1.26 $ft^2$ per kW |
| Capital Contingency Factor | 35% |

ENR = 3031

## TABLE I

## DIRECT FIRED THERMAL PROCESSING OPERATION AND MAINTENANCE COST FACTORS

| | |
|---|---|
| RDF Labor (Work Positions)(1) | (Number of furnace units) x 3 if RDF is fed, otherwise 0 |

Furnace Unit Labor (Work Positions)(1)

If furnace load (wet t/d) LE 1,300 = 5
If furnace load (wet t/d) GT 1,300, LE 3,900 = 10
If furnace load (wet t/d) GT 3,900, LE 6,500 = 15
If furnace load (wet t/d) GT 6,500, LE 8,667 = 20
If furnace load (wet t/d) GT 8,667, LE 10,833 = 25
If furnace load (wet t/d) GT 10,833, LE 13,000 = 30

| | |
|---|---|
| Fuel Cost | $0.40 per gallon (if used) |
| Electricity Cost and Credit | $0.03 per kWh |
| Maintenance Parts Cost | 3% x annual process and structural capital cost |

CPI = 183

(1) A work position entails a commitment of 8 man-hours per work position per day.

## TABLE J

## DIRECT FIRED THERMAL PROCESSING ELECTRICAL POWER USE AND PRODUCTION FACTORS

| | |
|---|---|
| Power Usage (kW) | Wet infeed (t/d) x 75 kW/wet ton |
| Power Produced (kW) | Weight of steam (600 psig - 800°F) produced/10.1 |
| Annual Energy Usage kWh/yr | Power usage x 24 x 365 |
| Annual Energy Produced kWh/yr | Power produced x 24 x 365 |

## TABLE K

## DIRECT FIRED THERMAL PROCESSING AIR EMISSION FACTORS

| | |
|---|---|
| Particulate Emission | 0.05 grains/DSCF stack gas |
| $NO_x$ | 0.596 lbs/dry ton stack gas plus 0.155 lbs $NO_x$ per MMBtu fuel oil burned. |
| $SO_2$ | 0.243 lbs/dry ton stack gas plus 0.100 lbs $SO_2$ per MMBtu fuel oil burned. |